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BIBLIOGRAPHIC DATA SHEET

1. Report No. APTD-0919

2.

3. Recipient's Accession No.

4. Title and Subtitle

Cost Nomographs of Selected Sulfur Dioxide Abatement Methods

5. Report Date January 1972

6.

7. Author(s)

8. Performing Organization Rept. No. HIT-508

9. Performing Organization Name and Address

Hittman Associates, Inc.
Columbia, Maryland

10. Project/Task/Work Unit No.

11. Contract/Grant No. EHSD-71-43

12. Sponsoring Organization Name and Address

Environmental Protection Agency
Office of Air Programs

13. Type of Report & Period Covered

14.

15. Supplementary Notes

16. Abstracts

The results are presented of a study which was undertaken to provide a basis for assessing the costs and effectiveness of alternative sulfur dioxide control systems for emissions of fossil fuel combustion from power generating facilities. The study attempts to provide a set of nomographs which would allow rapid and accurate assessment of these control alternatives. Sulfur emissions can be controlled by either stack gas control processes, desulfurization of oil or coal, or the burning of naturally-occurring low sulfur fuels. The economic aspects involved are the capital and operating costs of stack gas and desulfurization facilities and the incremental costs of purchasing and shipping low sulfur fuels. The most current technical and economic data were compiled and evaluated for six stack gas processes. The costs per ton of coal or per barrel of oil were determined given a variety of conditions. The cost of transporting bulk reactants or low sulfur fuels can be significant when studying control economics.

17. Key Words and Document Analysis. 17a. Descriptors

Air pollution
Electric power plants
Fossil fuels
Sulfur compounds
Air pollution control equipment
Desulfurization
Transportation
Cost estimates
Cost comparison

17b. Identifiers/Open-Ended Terms

Air pollution control

17c. COSATI Field/Group 13B

18. Availability Statement

Unlimited

19. Security Class (This Report) UNCLASSIFIED

20. Security Class (This Page) UNCLASSIFIED

21. No. of Pages ~~152~~ 148

22. Price $3.00

COST NOMOGRAPHS OF SELECTED SULFUR DIOXIDE ABATEMENT METHODS

HIT-508

January, 1972

Prepared Under
Contract No. EHSD-71-43
Environmental Protection Agency
Office of Air Programs

HITTMAN ASSOCIATES, INC.
COLUMBIA, MARYLAND

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I. INTRODUCTION

The continued economic growth of the United States requires ever-increasing supplies of electrical power. From past events, the power industry has needed to double its capacity each decade merely to remain even with the demand. This growth trend is further expected to continue through at least the end of this century. Nuclear power generating facilities will provide a substantial percentage of the required power through the next 30 years. However, fossil fuels will continue to be used in greater amounts until about the turn of the century.

A growing awareness on the part of government, industry, and the general population of the potential harmful effects on the environment and the ecology from continued industrial growth has spurred a nationwide program to control and to abate environmental impacts. As part of this program, large efforts and many millions of dollars have been expended to find ways in which the emissions of fossil fuel combustion from power generating facilities could be minimized. A variety of techniques for the control of effluents from power plants have been proposed, designed, or constructed.

In order to provide a basis for assessing the costs and effectiveness of alternative sulfur dioxide control systems, the Environmental Protection Agency (EPA) contracted with Hittman Associates, Inc., to perform a major study to provide a set of nomographs which would allow rapid and accurate assessment of these alternatives for control. This document contains the results of that study.

Sulfur emissions can be controlled by either stack gas control processes, desulfurization of oil or coal, or the burning of naturally-occurring low sulfur fuels. The economic aspects involved are the capital and operating costs of stack gas and desulfurization facilities and the incremental costs of purchasing and shipping low sulfur fuels.

The main emphasis was placed on the system and economic analysis of six stack gas processes. The most current technical and economic data were compiled and evaluated. The major cost parameters were identified and a method whereby the interrelation of these parameters could be depicted graphically was developed. The resulting nomographs allow the calculation of operating costs for any combination of plant size, fuel, sulfur content, load factor, fixed charge rate, additive cost, etc. Thus, the optimum control process can be selected for any given power plant.

In addition to stack gas control methods, the economic feasibility of coal and oil desulfurization was studied. The costs per ton of coal or per barrel of oil were determined given a variety of conditions. Nomographs were developed which allow one to determine desulfurization costs and to compare these costs with those of the stack gas removal processes.

The cost of transporting bulk reactants or low sulfur fuels can be significant when studying control economics. For this reason, a nomograph was developed which allows the determination of transportation costs. All fuel forms (coal, gas, and oil) and modes of transportation were considered.

II. NOMOGRAPHS

The major purpose of the nomographs presented in this section is the calculation of SO_2 abatement method costs. The nomographs dealing with stack gas SO_2 removal systems calculate the fixed and variable costs of the system when operated under known conditions. Typical inputs are the plant size, load factor, heat rate, fuel, and the sulfur content in fuel. Annual costs, as well as cost per kilowatt-hour, can be obtained.

Nomographs appearing in this section include the following: (Page numbers in parentheses refer to the separate nomograph section at the end of this chapter.)

- Plant size and Fuel Consumption Nomographs (pages 1-4)
- Dry Limestone Nomographs (pages 5-9)
- Wet Limestone Nomographs (pages 10-13)
- Magnesium Oxide Nomographs (pages 14-16)
- Catalytic Oxide Nomographs (pages 17-19)
- Modified Chamber Nomographs (pages 20-22)
- Molten Carbonate Nomographs (pages 23-26)
- Operating Cost Nomographs (pages 27-29)
- Desulfurization and Transportation Nomographs (pages 30-35)

To allow convenient use of the process nomographs, work sheets have been included. These sheets can be reproduced to provide any number of additional forms. They provide the proper format for the step-by-step utilization of the nomographs. Table II-1 refers to nomographs on pages 1-4 in which the basic computations necessary for all processes are performed. Table II-2 refers to the process nomographs on pages 5-26. Table II-3 is included to provide a first estimate for all required input variables. A low, typical, and high value for each parameter is included to bracket the expected range. Appendix A tabulates example cases for each process. These examples will assist the individual's first attempts at using the nomographs.

Utilizing the nomograph on page 27, the annual quantities of sulfur removed and sulfur released upon combustion may be determined as a function of energy utilization, sulfur content of the fuel and the process SO_2 removal efficiency. The nomographs on pages 28 and 29 provide alignment charts used to convert annual operating costs into mills per kilowatt-hour, dollars

TABLE II-1. INITIAL CALCULATIONS

Page	Inputs		Outputs	
1	A. Heat Rate (Optional) 1. Plant overall electrical efficiency, %		2. Plant net heat rate, Btu/Kw-hr	
1	B. Capital Recovery Factor 3. Interest rate % 4. Years amortized 6. Recurring costs, %/yr		5. Capital recovery factor, %/yr 7. Fixed charge rate, %/yr (add 5 and 6)	
2	C. Energy Utilization see 2 above plus . . . 8. Net generating capacity, Mw 9. Plant load factor, %		10. Annual energy production, 10^9 Kw-hr/yr 11. Energy utilization, 10^{12} Btu/yr	
3	D. Fuel Cost and Consumption see 11 above plus . . . 12. Heat value 14. Fuel cost (optional)		13. Annual fuel consumption 15. Energy cost (optional)¢/10^6 Btu	
4	E. Sulfur Content Unit Conversion see 12 above plus . . . 16. Weight percent of sulfur in fuel		17. Sulfur content, lb/10^6 Btu	
4	F. Flue Gas Flow Rate see 2 and 8 above		18. Flue gas flow rate, 10^6 scfm	

TABLE II-2. PROCESS CALCULATIONS

Page	Inputs		Outputs	
	Dry Limestone Process			
5	A. Efficiency 19. Stoichiometric ratio		20. Sulfur dioxide removal efficiency, %	
6	B. Limestone Use Factor see 17 and 19 above plus . . . 21. Limestone purity, wt-% of carbonate		22. Limestone use factor/lb/10^6 Btu	
7	C. Fixed Cost see 7, 18, and 22 above . . .		23. Fixed capital cost, $10^6\$$ 24. Fixed cost, $10^6\$$/yr	
8	D. Other Operating Costs see 11 and 22 above . . .		25. Other operating costs, $10^6\$$/yr	
8	E. Grinding Correction (Optional) 26. Limestone particle size, μ		27. Incremental cost of grinding, $\$$/ton	
9	F. Additive Cost see 11 and 22 above plus . . . 28. Total reactant cost, $\$$/ton-add delivered limestone cost, incremental cost of grinding (see 27), and disposal cost		29. Annual additive cost, $10^6\$$/yr	
	Dry Limestone Total Operating Cost, $10^6\$$/yr		Add 24, 25, and 29	
	Wet Limestone Process			
10	A. Efficiency Assumed constant at 85%			
10	B. Limestone Use Factor see 17 above plus . . . 30. Stoichiometric ratio 31. Limestone purity, wt-% of carbonate		32. Limestone use factor, lb/10^6 Btu	
11	C. Fixed cost see 7, 18, and 32 above		33. Fixed capital cost, $10^6\$$ 34. Fixed cost, $10^6\$$/yr	
12	D. Other Operating Costs see 11 and 32 above		35. Other operating costs, $10^6\$$/yr	
12	E. Reheat Correction (Optional) see 18 above plus . . . 36. Reheat temperature, °F		37. Credit, $10^6\$$/yr	
13	F. Additive Cost see 11 and 32 above plus . . . 38. Total reactant cost, $\$$/ton-add delivered limestone cost and disposal cost		39. Annual additive cost, $10^6\$$/yr	
	Wet Limestone Total Operating Cost, $10^6\$$/yr		Add 34, 35, and 39, subtract 37	
	Magnesium Oxide Process			
14	A. Fixed Cost see 7, 17, and 18		40. Fixed capital cost, $10^6\$$ 41. Fixed cost, $10^6\$$/yr	
15	B. By-Product Acid Credit see 11 and 17 above plus . . . 43. $\$$/ton of 98.5% H_2SO_4		42. Sulfur removed, 10^3 ton/yr 44. Acid credit, $10^6\$$/yr	
16	C. Raw Material Cost see 11 above plus . . . 45. $\$$/ton of magnesia		46. Annual MgO cost, $10^6\$$/yr	
16	D. Other Operating Costs see 11 and 17 above		47. Other operating costs, $10^6\$$/yr	
	Magnesium Oxide Total Operating Cost, $10^6\$$/yr		Add 41, 46, and 47, subtract 44	

TABLE II-2. PROCESS CALCULATIONS (Continued)

Page	Inputs		Outputs	
	Catalytic Oxidation Process			
17	A. Fixed Cost see 7, 17, and 18 above		48. Fixed capital cost, $10^6\$$ 49. Fixed cost, $10^6\$/\text{yr}$	
18	B. By-Product Acid Credit see 17 and 18 above plus . . . 51. \$/ton of acid		50. Sulfur removed, 10^3 tons/yr 52. Acid credit, $10^6\$/\text{yr}$	
19	C. Raw Material Cost see 11 above plus . . . 53. % ash in fuel 55. V_2O_5 cost (\$/ft^3)		54. Quantity of V_2O_5 catalyst, ft^3/yr 56. Annual V_2O_5 catalyst cost, $10^6\$/\text{yr}$	
19	D. Other Operating Costs see 11 above		57. Other operating costs, $10^6\$/\text{yr}$	
	Catalytic Oxidation Total Operating Cost, $10^6\$/\text{yr}$		Add 49, 56, and 57, subtract 52	
	TYCO Modified Chamber Process			
20	A. Fixed Cost see 7 and 18 above		58. Fixed capital cost, $10^6\$$ 59. Fixed cost, $10^6\$/\text{yr}$	
21	B. Raw Material Cost see 11 above plus . . . 60. Charcoal catalyst cost, \$/ft^3		61. Annual catalyst cost, $10^6\$/\text{yr}$	
21	C. Other Operating Costs see 11 above		62. Other operating costs, $10^6\$/\text{yr}$	
22	D. By-Product Acid Credits see 11 and 17 65. \$/ton of 80% H_2SO_4 67. NO_x volume content, ppm 69. \$/ton of 60% HNO_3		63. Sulfur removed, 10^3 ton/yr 64. 80% H_2SO_4, 10^3 ton/yr 66. By-product H_2SO_4, $10^6\$/\text{yr}$ 68. 60% HNO_3, 10^3 ton/yr 70. By-product HNO_3, $10^6\$/\text{yr}$	
	TYCO Modified Chamber Process, $10^6\$/\text{yr}$		Add 59, 61, and 62, subtract 66 and 70	
	Molten Carbonate Process			
23	A. Fixed Cost see 7, 11, and 17 above		71. Fixed capital cost, $10^6\$$ 72. Fixed cost, $10^6\$/\text{yr}$	
24	B. Other Operating Costs see 11 and 17 above		73. Other operating costs, $10^6\$/\text{yr}$	
24	C. By-Product Sulfur Credit see 11 and 17 above plus . . . 75. \$/ton of sulfur		74. Sulfur removed, 10^3 ton/yr 76. By-product sulfur credit, $10^6\$/\text{yr}$	
25 & 26	D. Raw Material Costs see 11 and 74 above plus . . . 77. \$/ton of coke 79. Percent of ash in fuel 82. Percent of chlorine in fuel		78. Annual coke cost, $10^6\$/\text{yr}$ (1) 80. M_2CO_3 Replacement cost, $10^6\$/\text{yr}$ (2) 81. M_2CO_3 Replacement cost, $10^6\$/\text{yr}$ (3) 83. M_2CO_3 Replacement cost, $10^6\$/\text{yr}$ (4)	
	Molten Carbonate Total Operating Cost, $10^6\$/\text{yr}$		Add 72, 73, 78, 80, 81, and 83, subtract 76	

TABLE II-3. RANGES OF INPUT PARAMETERS

Nomograph	Page No.	Input Identification	Low	Typical	High
Initial Calculations	1	1	20	30	40
		3	3	6	9
		4	5	10-15	30
		6	1	2-3	4
	2	8	100	500	1,000
		9	20	50	75
	3	12 (coal only)	6000	12,000	14,000
		14 (coal only, \$10/ton)	0.3	0.6	1.0
	4	16 (coal only)	2	3.5	5.0
Dry Limestone	5	19	1	2	3
	6	21	80	90	95
	9	28 Limestone	2	4	7
		Grinding (size dependent - see output 27)			
		Disposal	0.4	0.5	1.0
		Total	2.4	4.5	8.0
Wet Limestone	10	30	1	1.3	1.5
		31	80	90	95
	12	36	150	250	275
	13	38 (see 28)			
Magnesium Oxide	15	43	5	12	15
	16	45	40	50	80
Catalytic Oxidation	18	51	1	5	10
	19	53	1	10	15
		55	30	38	50
TYCO Modified Chamber	21	60	10	14	20
	22	65	1	5	10
		67	250	400	1,200
		69	10	15	35
Molten Carbonate	24	75	5	20	25
	25	77	5	8	20
		79	1	10	15
	26	82	0.01	0.05	0.5

per unit of fuel consumption, and dollars per ton of sulfur removed.

Pages 30-32 calculate the cost of fuel desulfurization. Examples of the use of these nomographs, as well as a technical discussion of fuel desulfurization, can be found in Section IV.E. Pages 33-35 include nomographs which estimate the cost of transporting oil, gas, coal, and solid reactants.

A cost sensitivity analysis was performed utilizing the nomographs. Major parameters such as plant size, load factor, sulfur content, and additive costs were varied and the effect on total operating cost determined. The results of the analysis are presented in Section III.

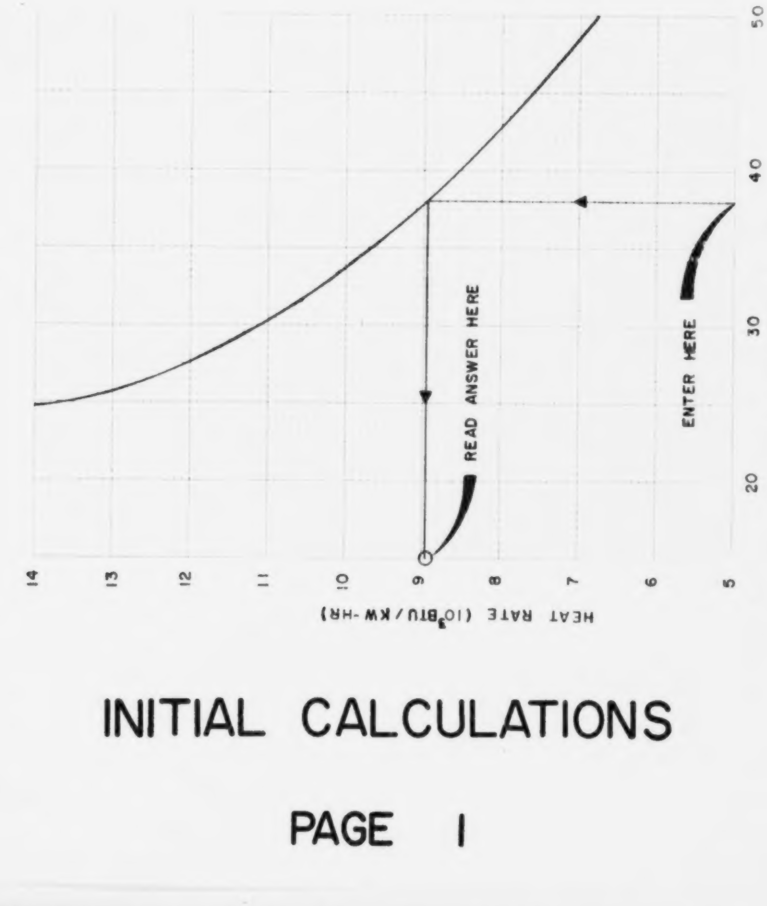
NOTICE

The nomographs which follow are based on the best currently available economic data. It is expected, however, that these data will change as process development progresses. Prototype tests currently being conducted or in the planning stages are expected to make significant contributions to the economic data base. Improvements in, or modifications to, system components are also expected to affect economics.

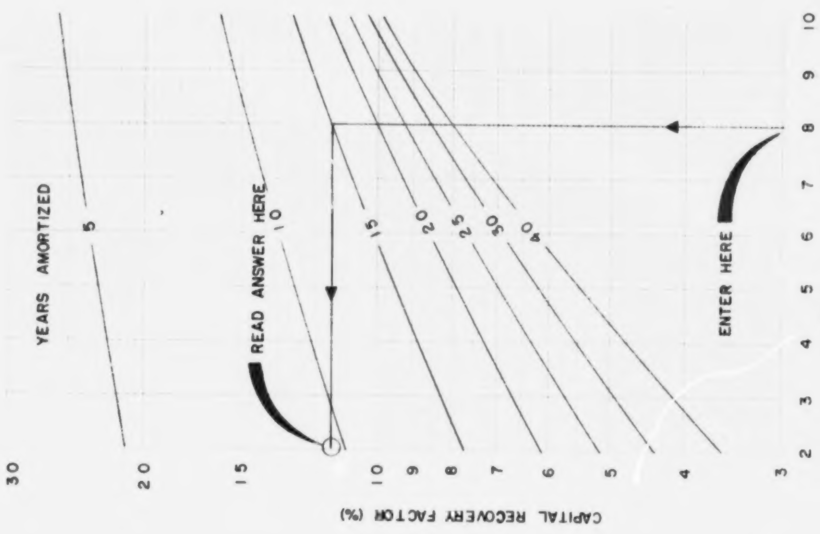
Economic comparisons of processes are meaningful only if they are at the same stage of development. Comparison of a fully developed process with one not yet in the prototype stage may not be accurate since the economics of the least developed process are based on assumptions which have not been thoroughly tested. Section IV. D discusses development status of the systems.

Many secondary economic variables have been held constant at typical or maximum values. This was required to simplify the nomographs. Section IV. C of this report discusses these variables in detail.

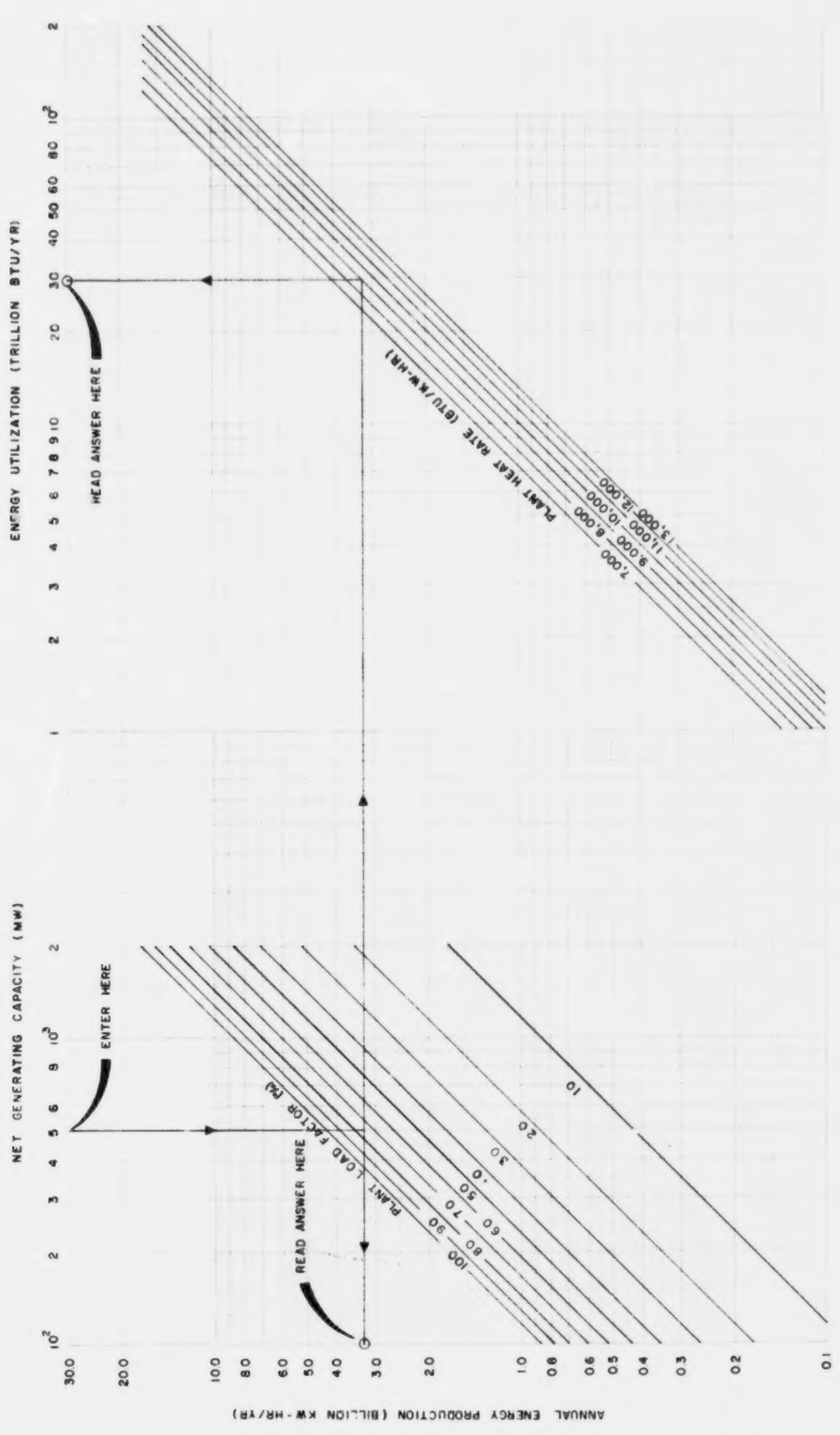
INITIAL CALCULATIONS



A-HEAT RATE



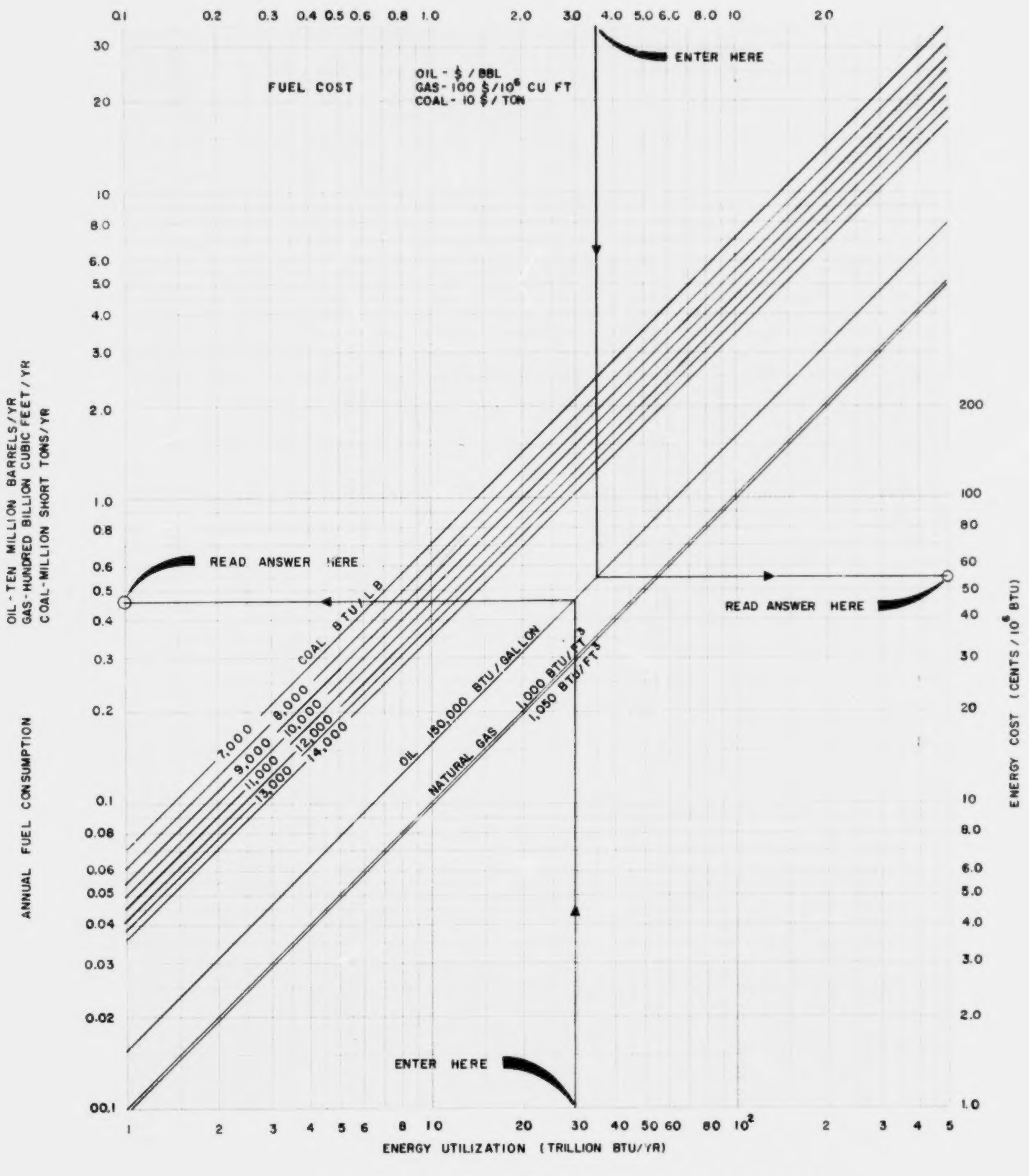
B-CAPITAL RECOVERY FACTOR



C-ENERGY UTILIZATION

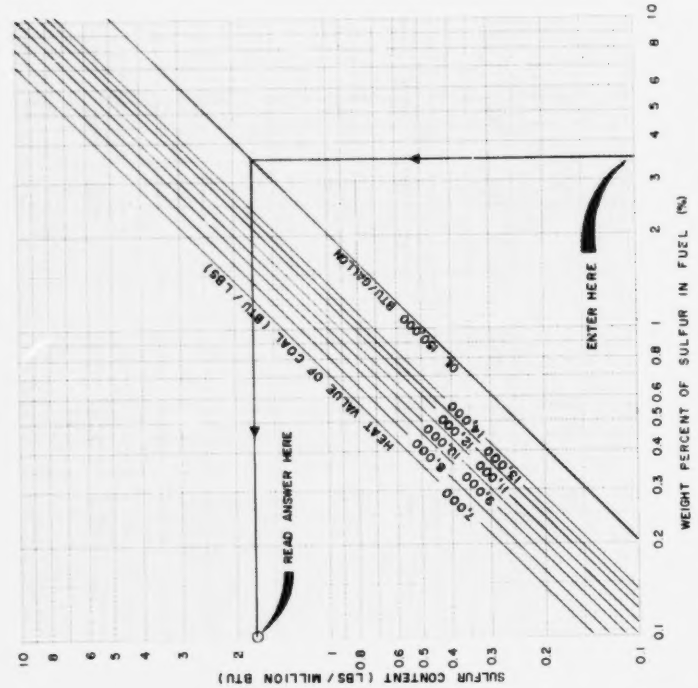
INITIAL CALCULATIONS (CONTINUED)

PAGE 2

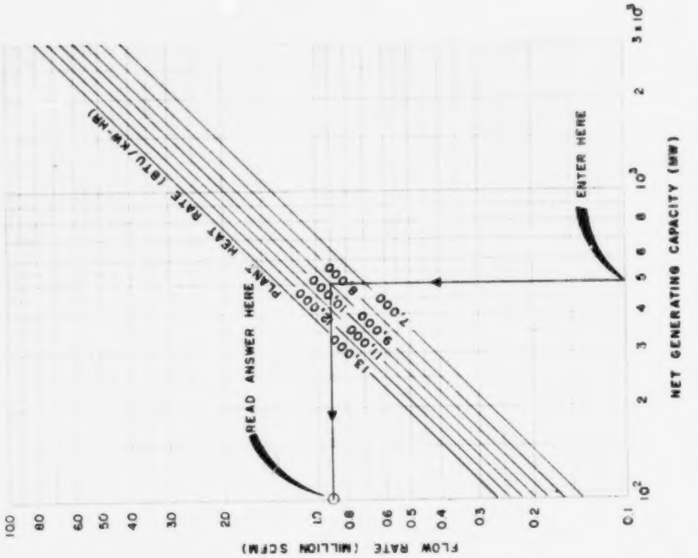


D-FUEL COST & CONSUMPTION

INITIAL CALCULATIONS (CONTINUED)



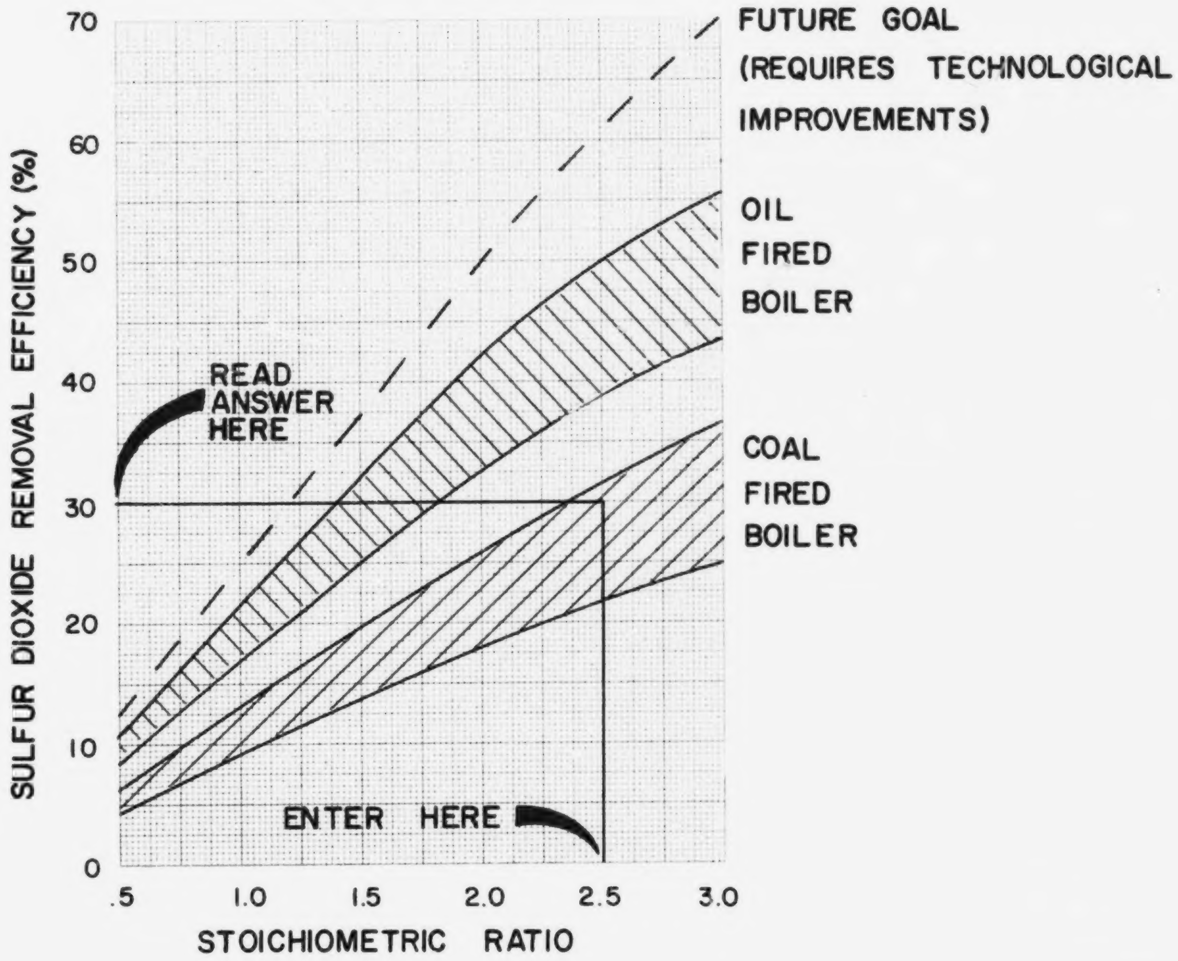
E - SULFUR CONTENT UNIT CONVERSION



F - FLUE GAS FLOW RATE

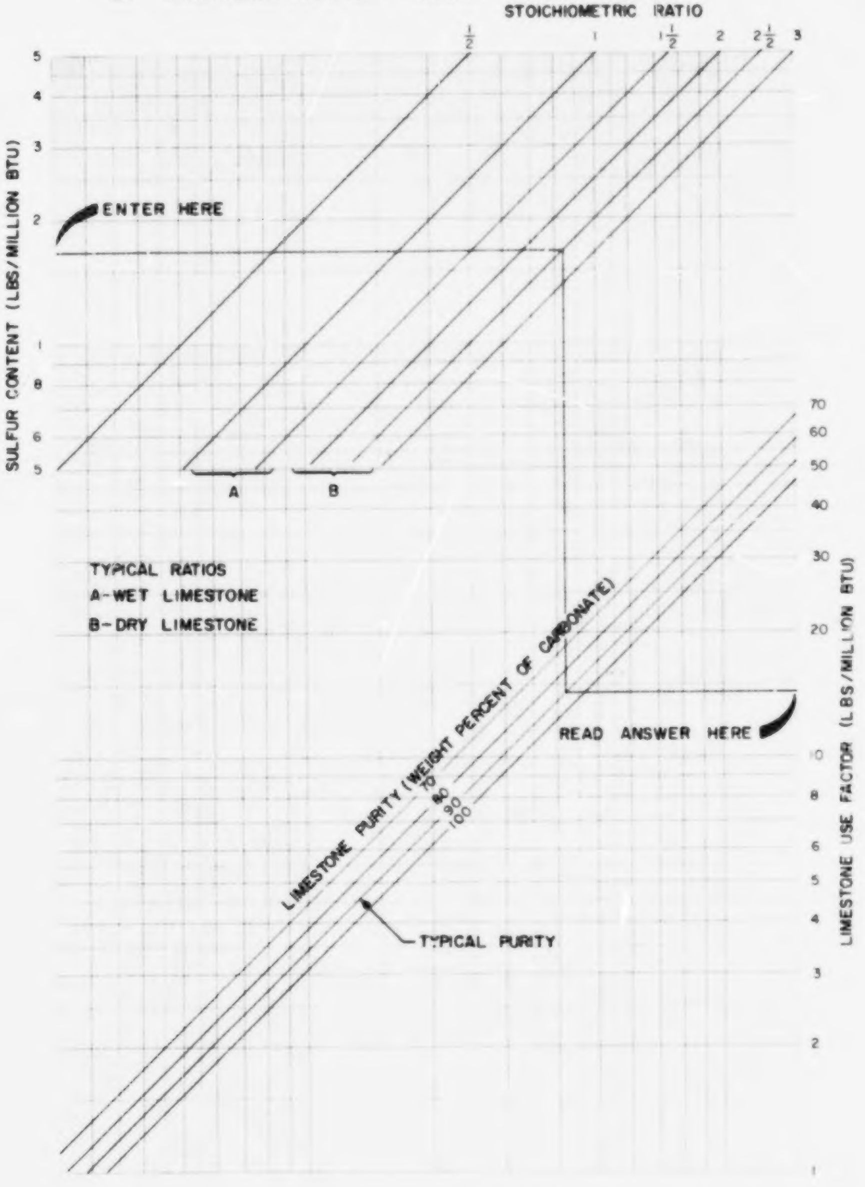
INITIAL CALCULATIONS (CONTINUED)

A-EFFICIENCY



DRY LIMESTONE

B-LIMESTONE USE FACTOR



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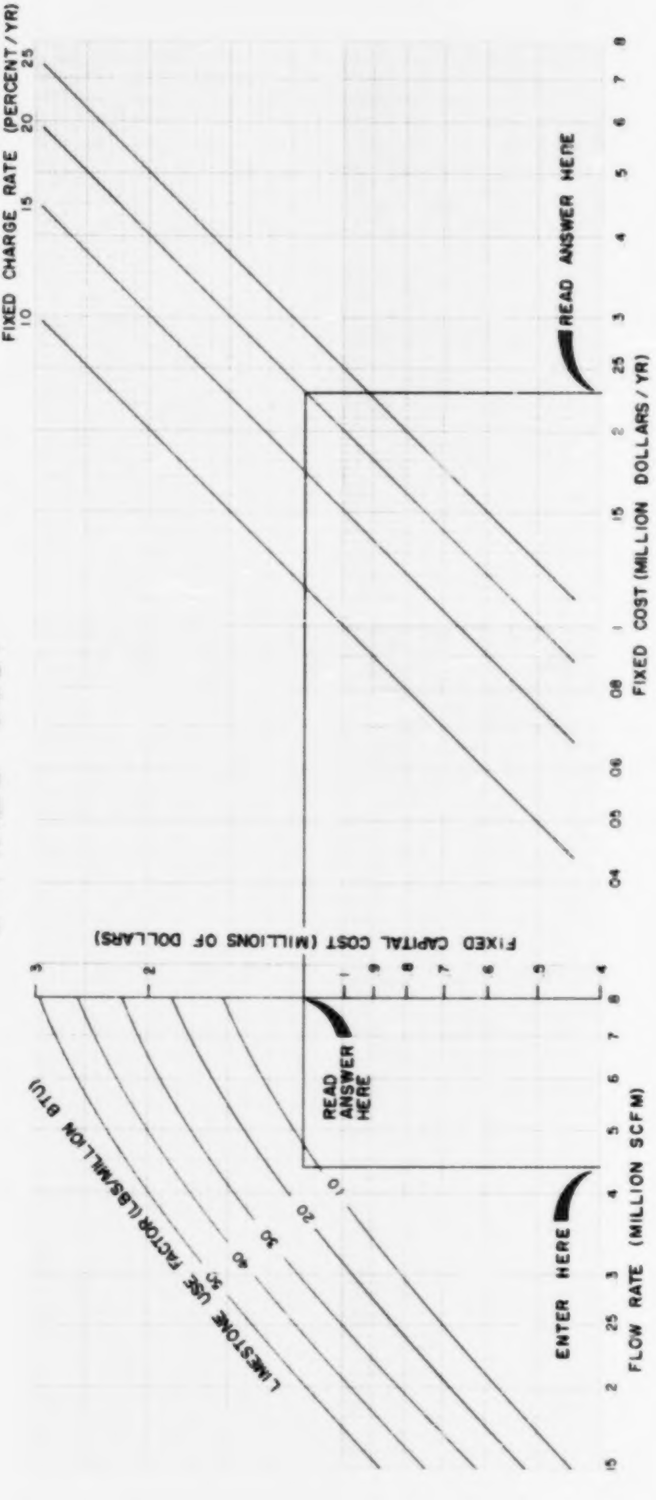
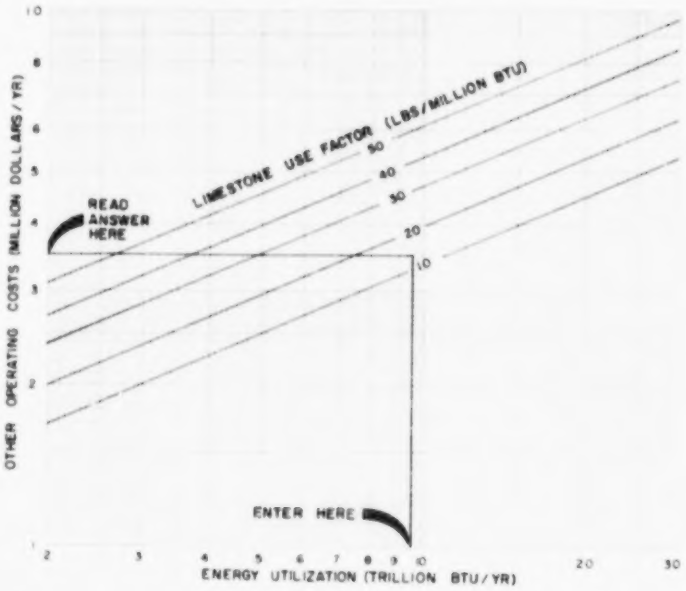
C-F!XED COST

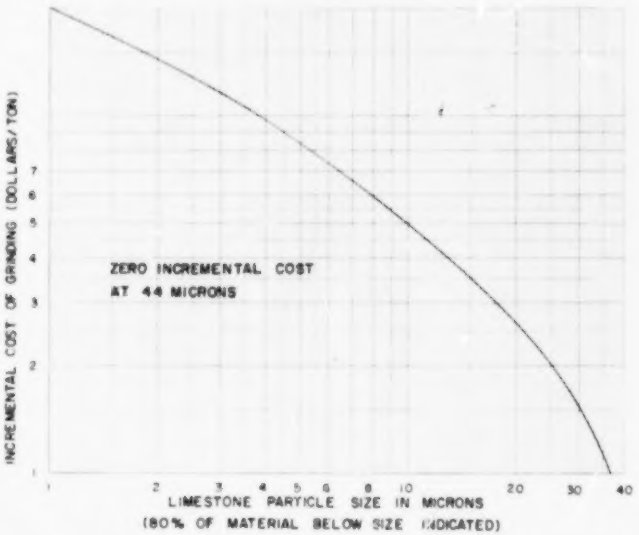


DRY LIMESTONE (CONTINUED)

D-OTHER OPERATING COSTS

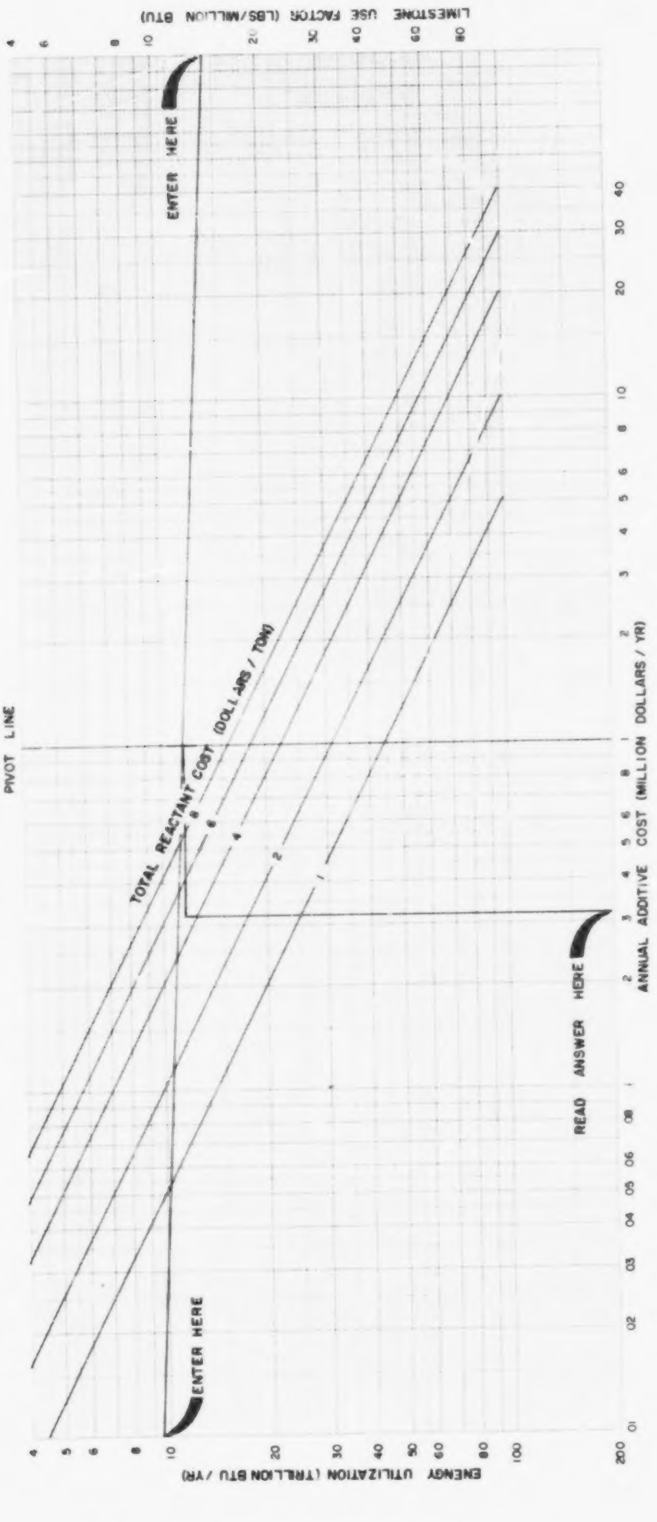


E-GRINDING CORRECTION



DRY LIMESTONE (CONTINUED)

F-ADDITIVE COST

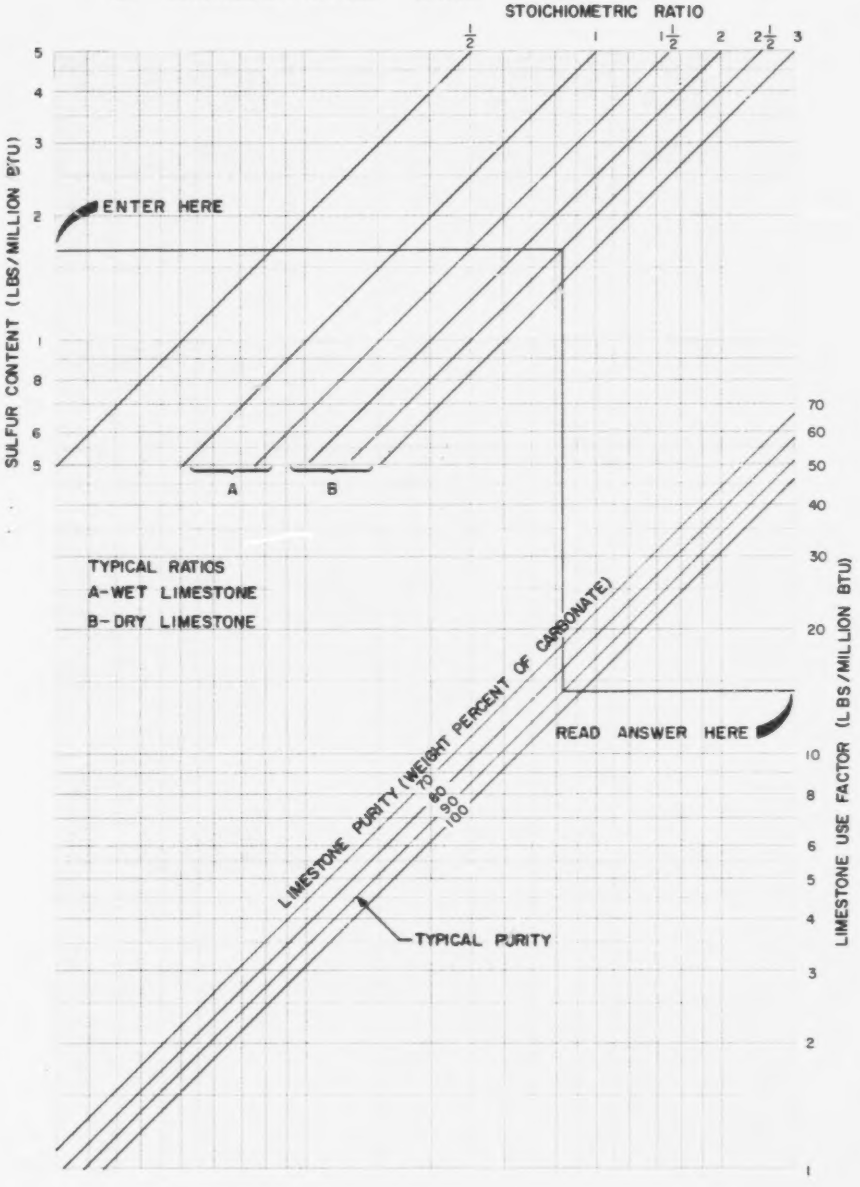


DRY LIMESTONE (CONTINUED)

A - EFFICIENCY

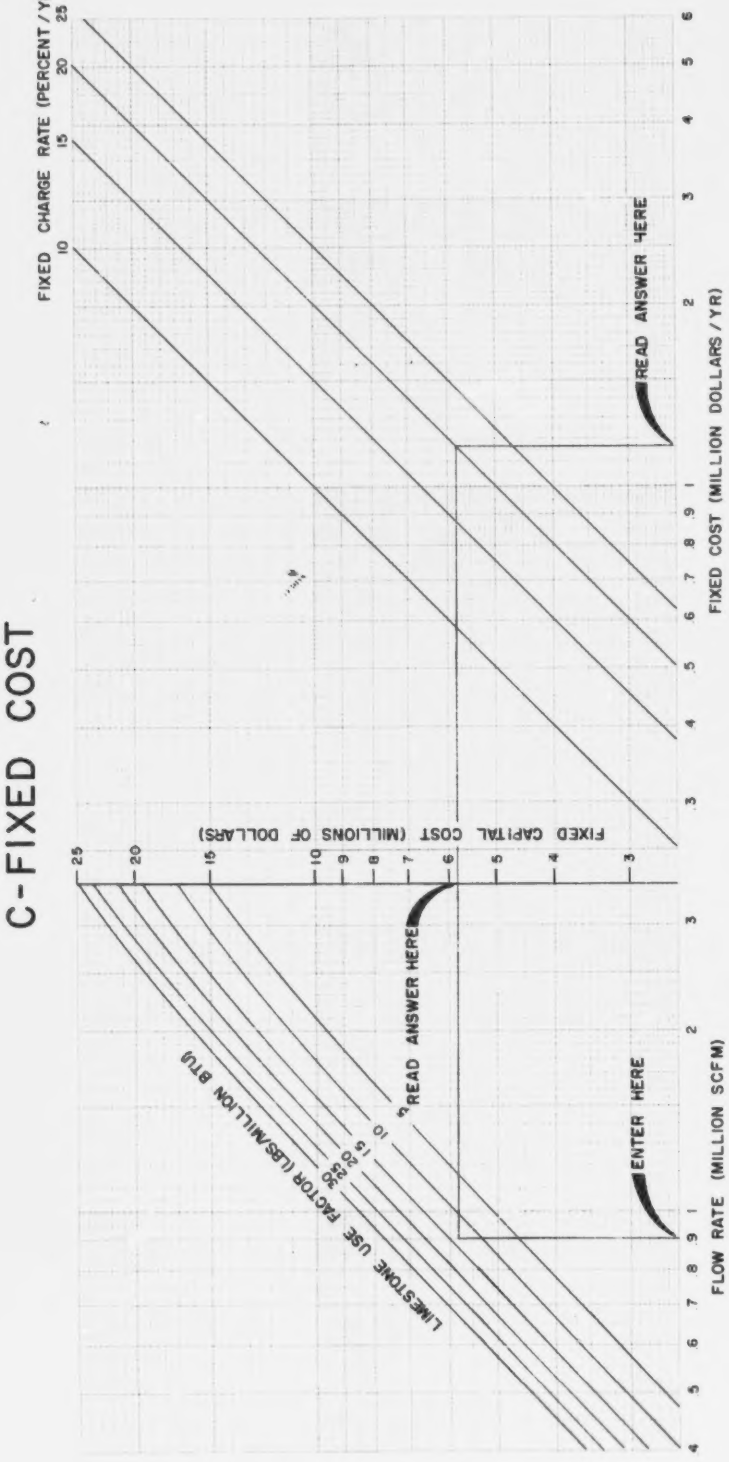
ASSUMED CONSTANT
AT 85 PERCENT

B - LIMESTONE USE FACTOR



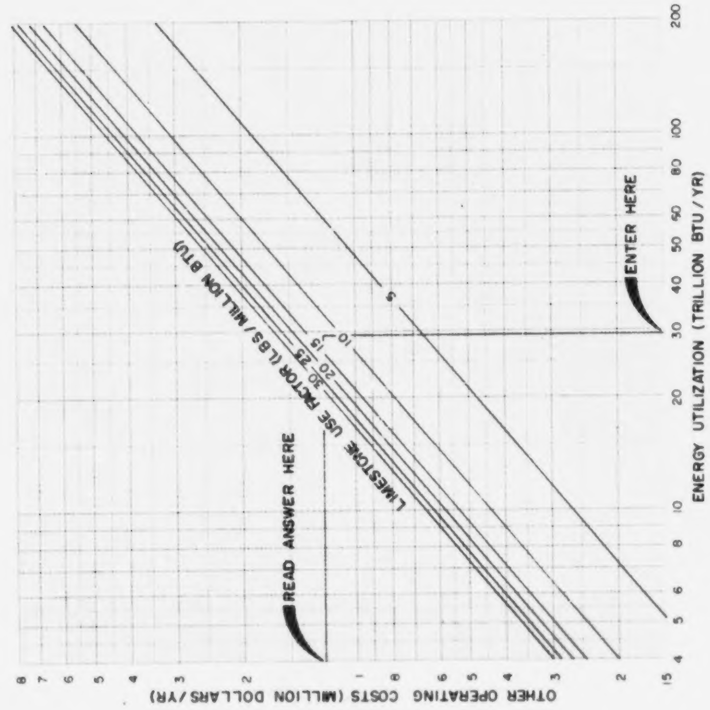
WET LIMESTONE

C-FIXED COST



WET LIMESTONE (CONTINUED)

D-OTHER OPERATING COSTS



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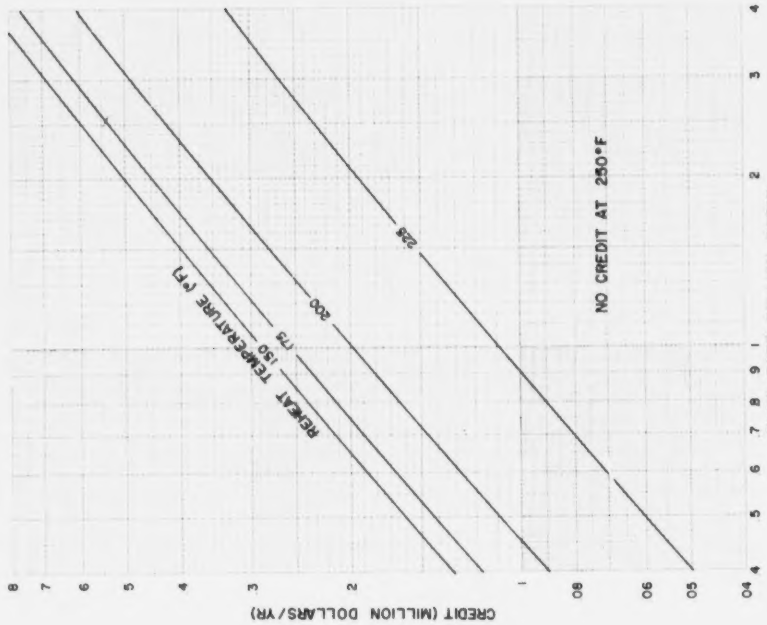
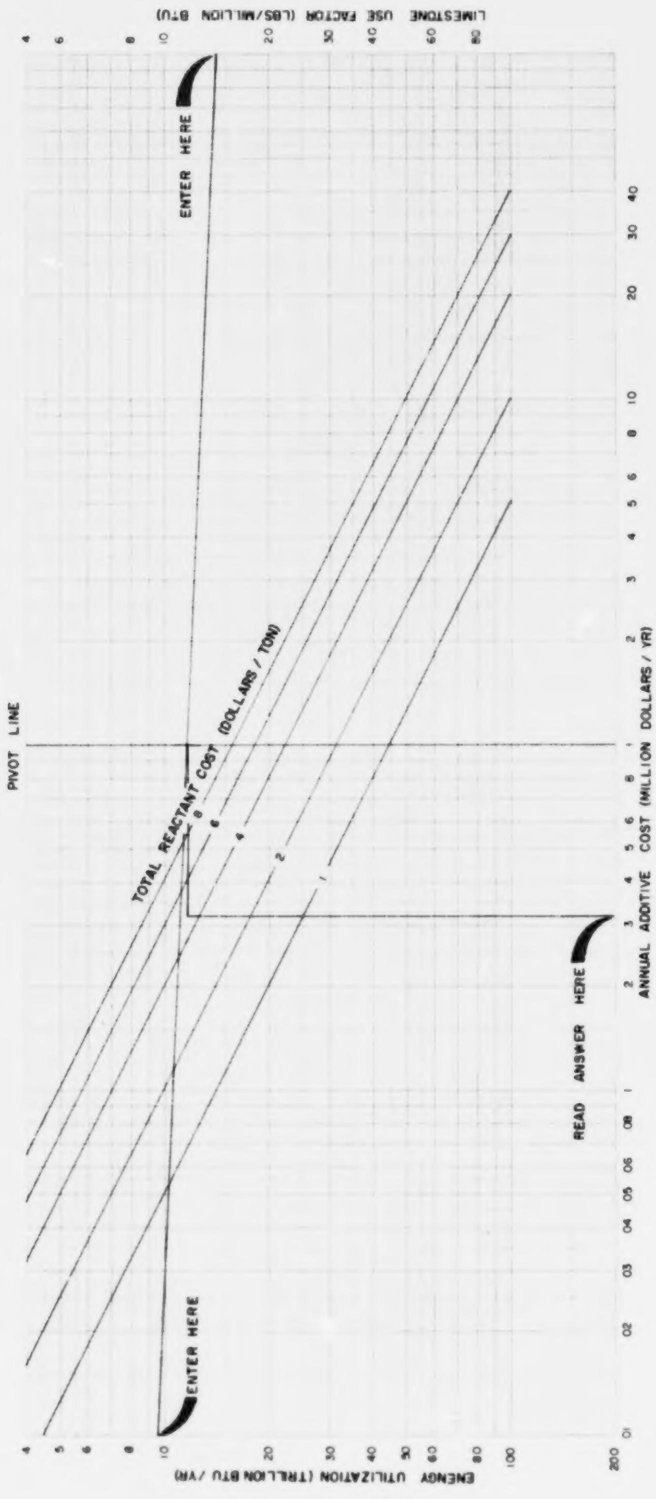


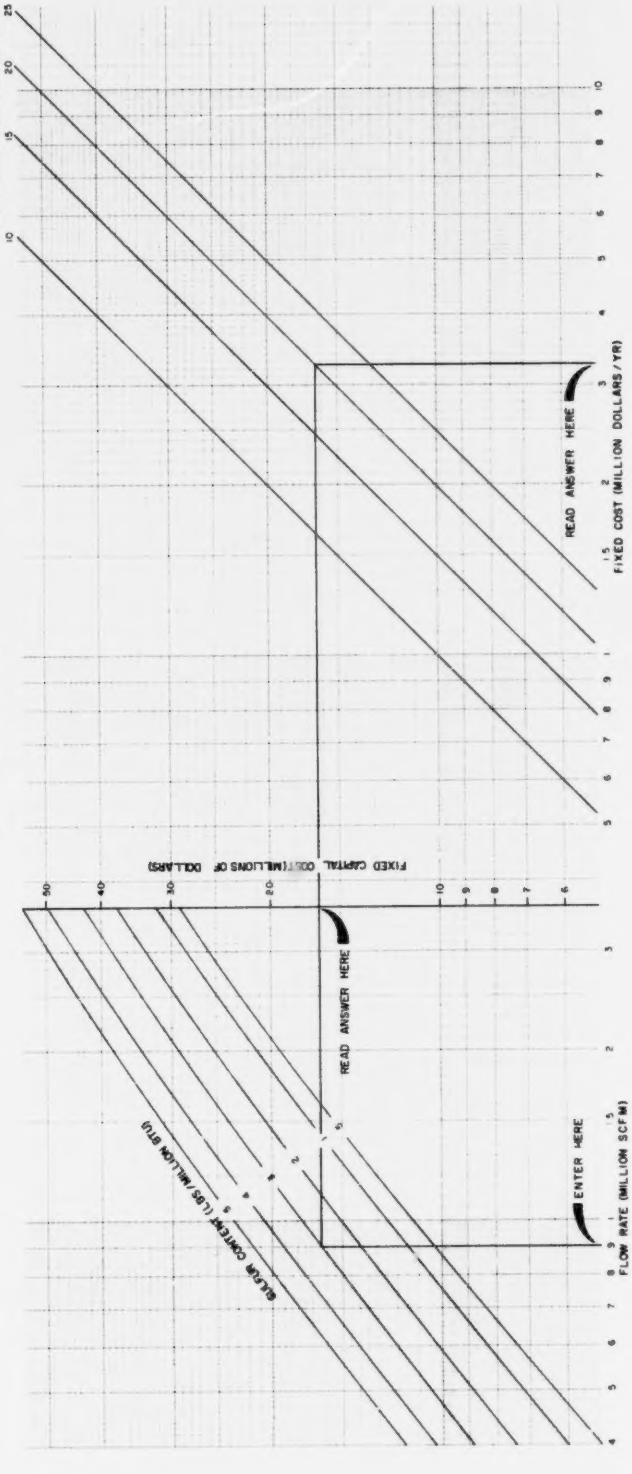
WET LIMESTONE (CONTINUED)

F-ADDITIVE COST



WET LIMESTONE (CONTINUED)

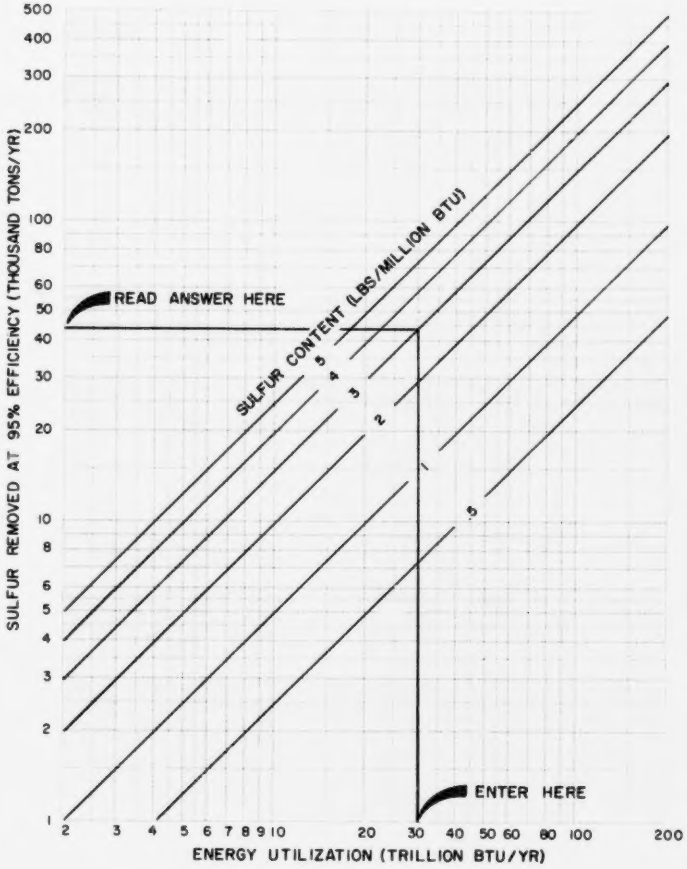
A-FIXED COST

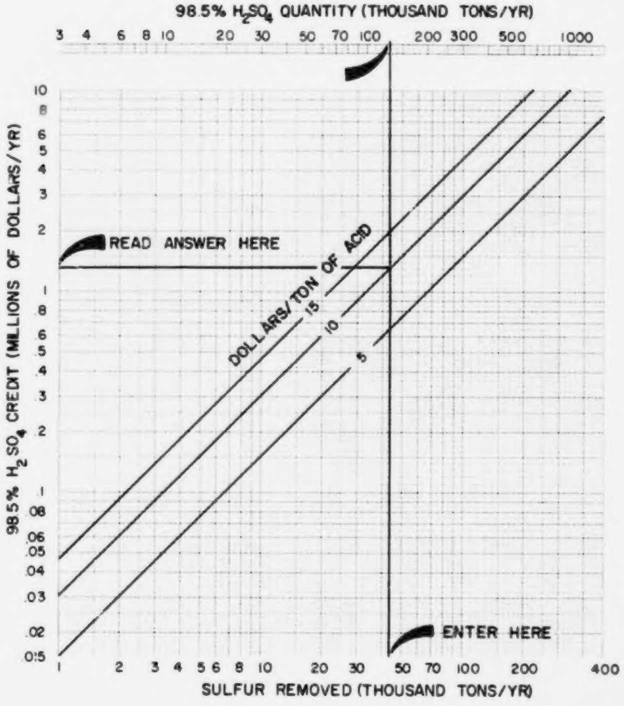


MAGNESIUM OXIDE

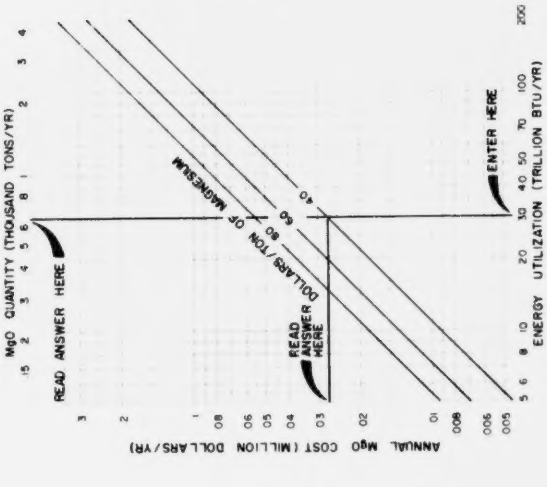
B-BY-PRODUCT ACID CREDIT

98.5 % H_2SO_4

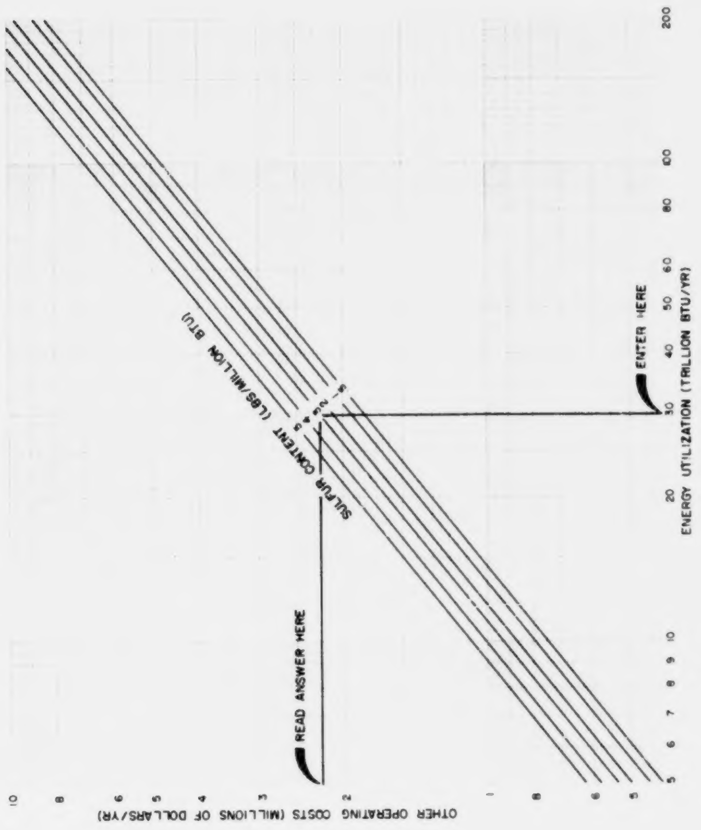




C-RAW MATERIAL COST MAGNESIA MAKEUP

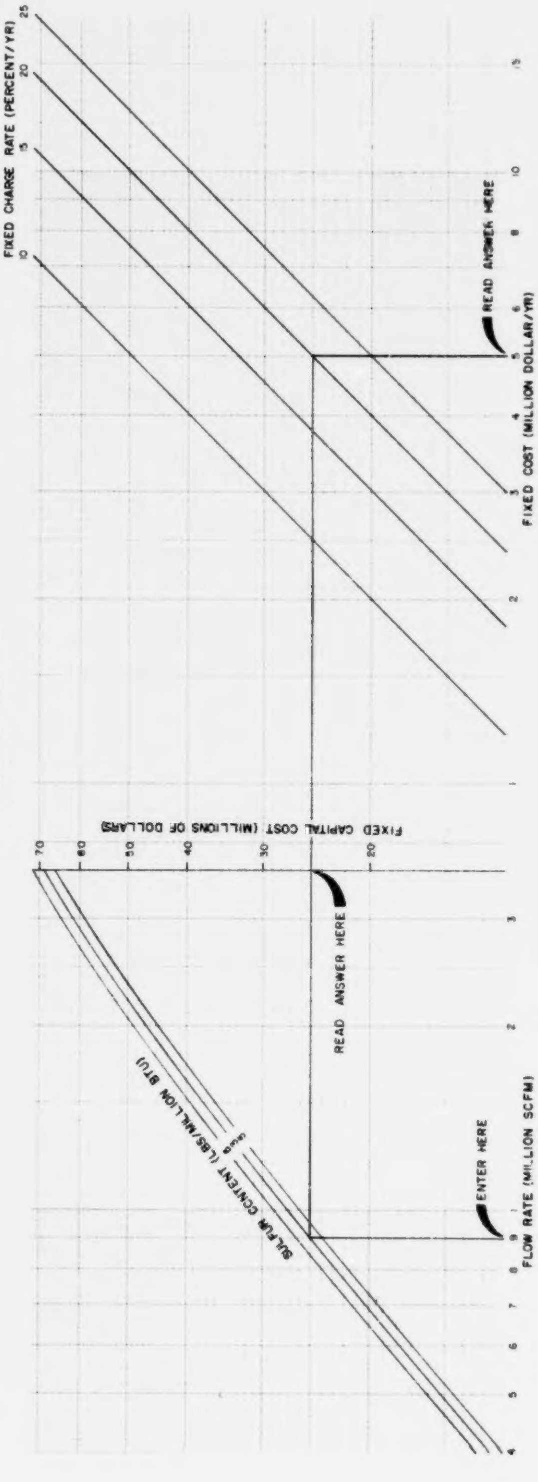


D-OTHER OPERATING COSTS



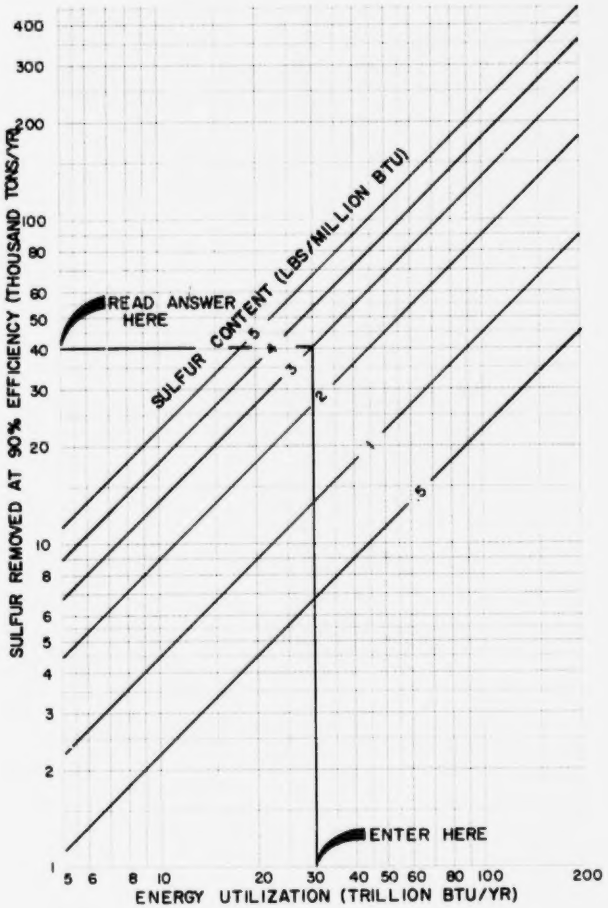
MAGNESIUM OXIDE (CONTINUED)

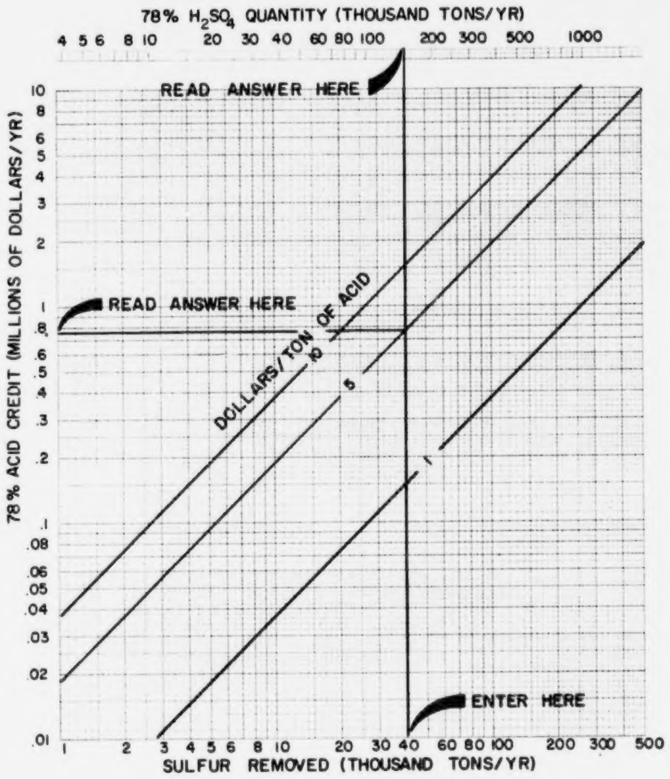
A-FIXED COST



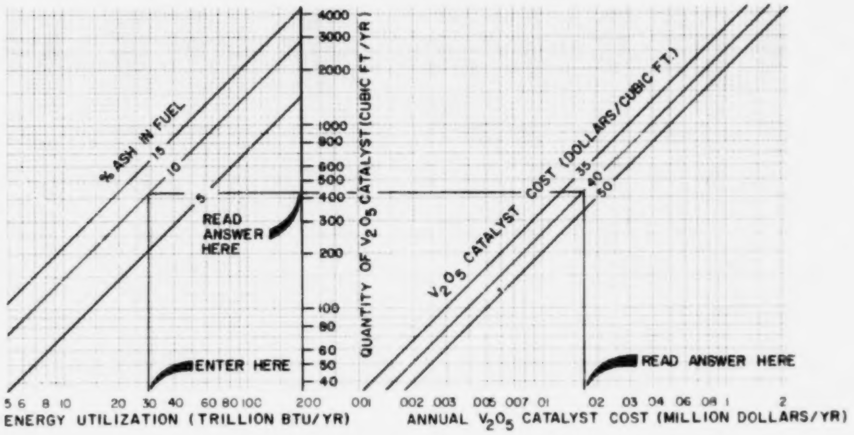
CATALYTIC OXIDATION

B-BY-PRODUCT ACID CREDIT
78% H_2SO_4





C-RAW MATERIAL COST
V_2O_5 CATALYST



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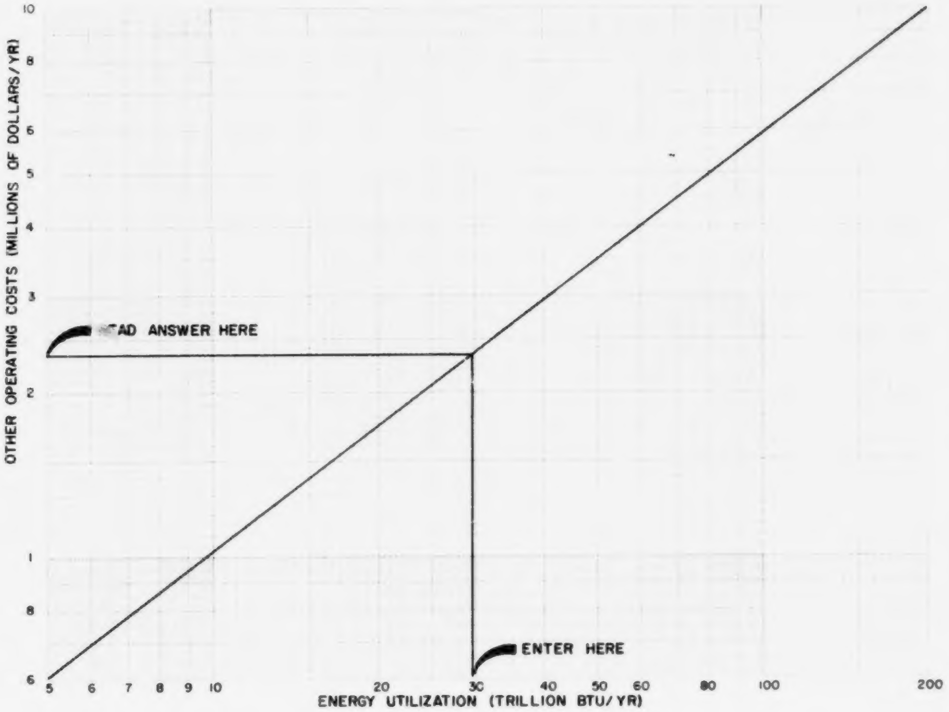
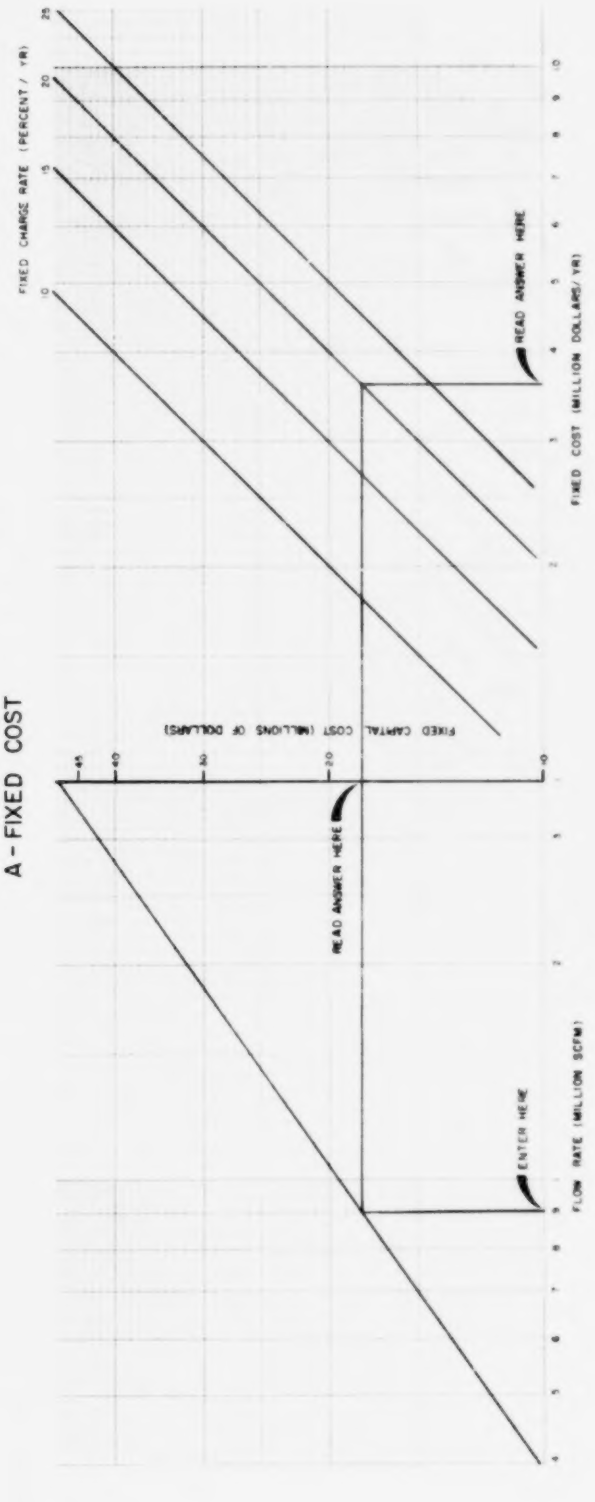
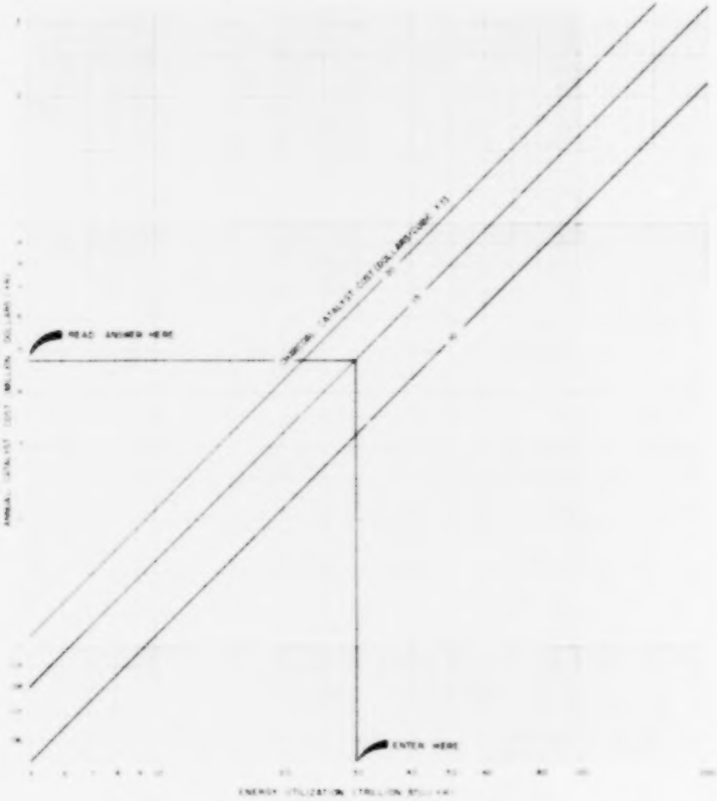


CATALYTIC OXIDATION (CONTINUED)

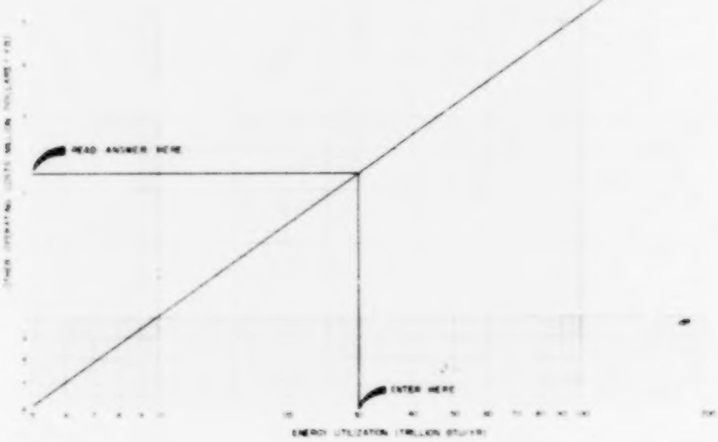


TYCO MODIFIED CHAMBER

B-RAW MATERIAL COST



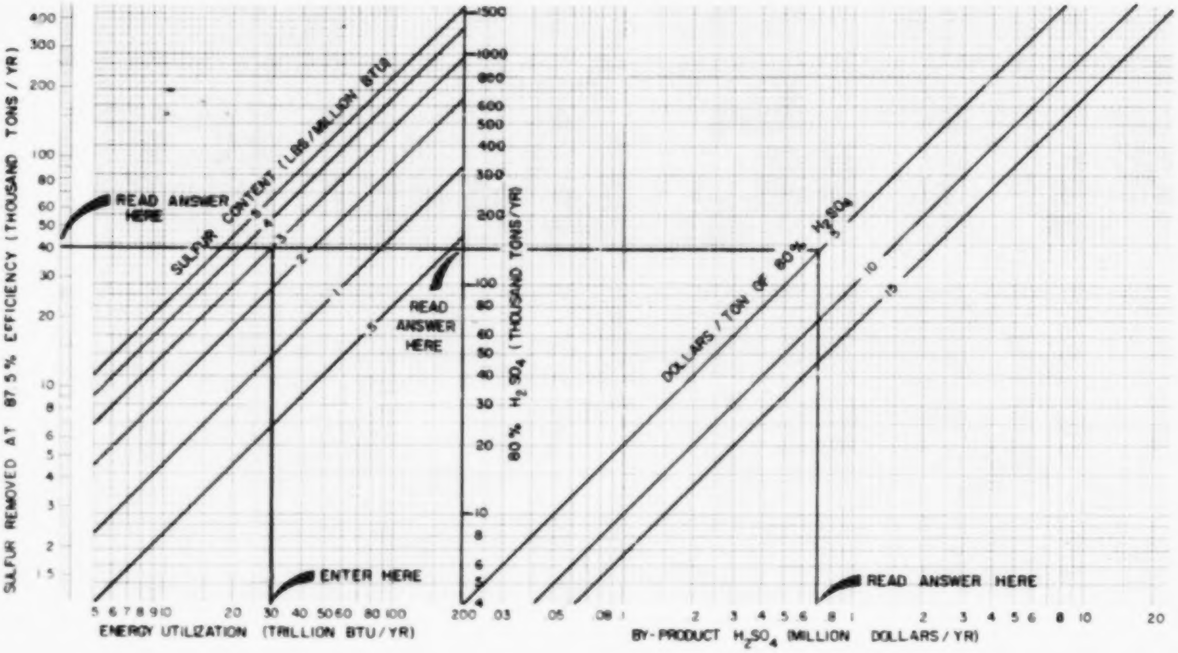
C-OTHER OPERATING COSTS



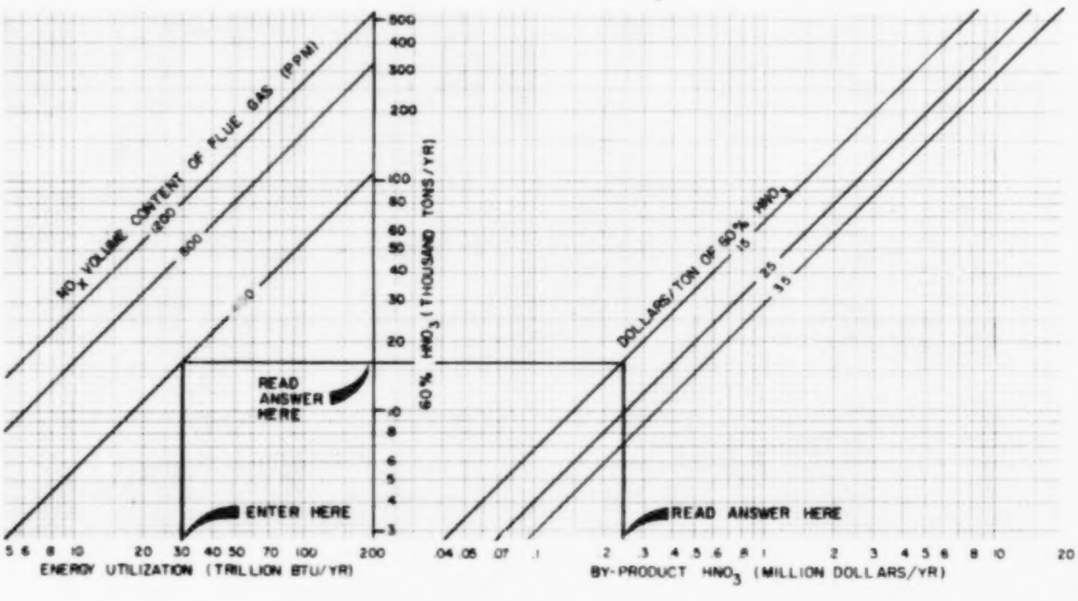
TYCO MODIFIED CHAMBER (CONTINUED)

D-BY-PRODUCT ACID CREDITS

1. 80% H_2SO_4

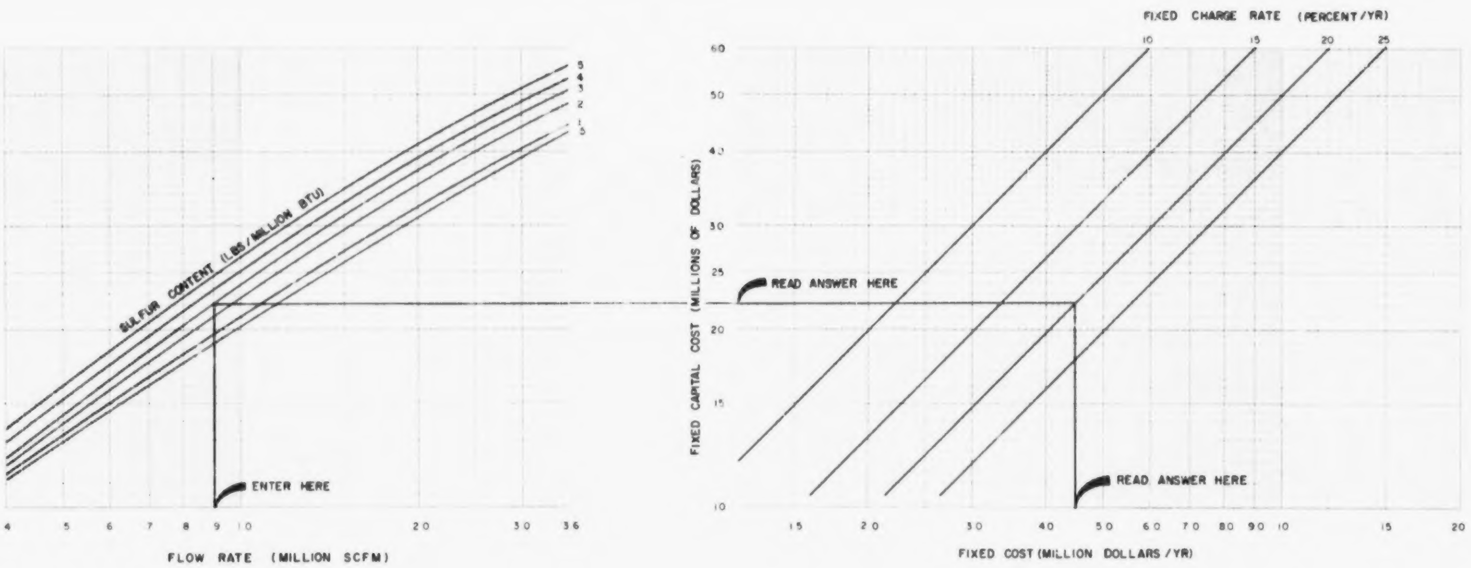


2. 60% HNO_3



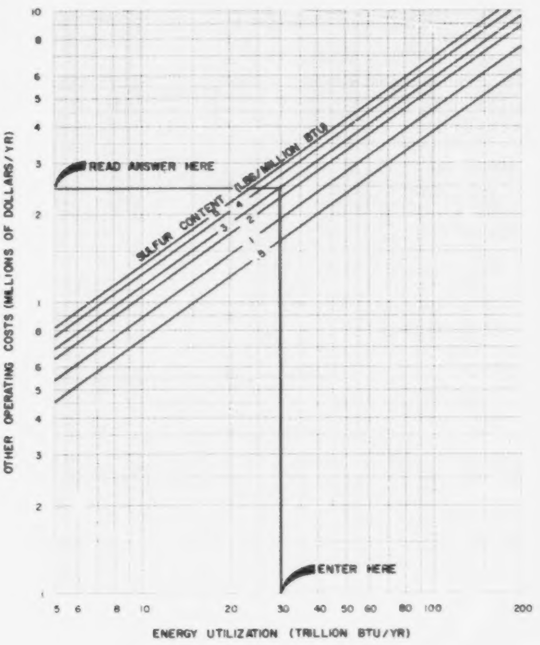
TYCO MODIFIED CHAMBER (CONTINUED)

A-FIXED COST

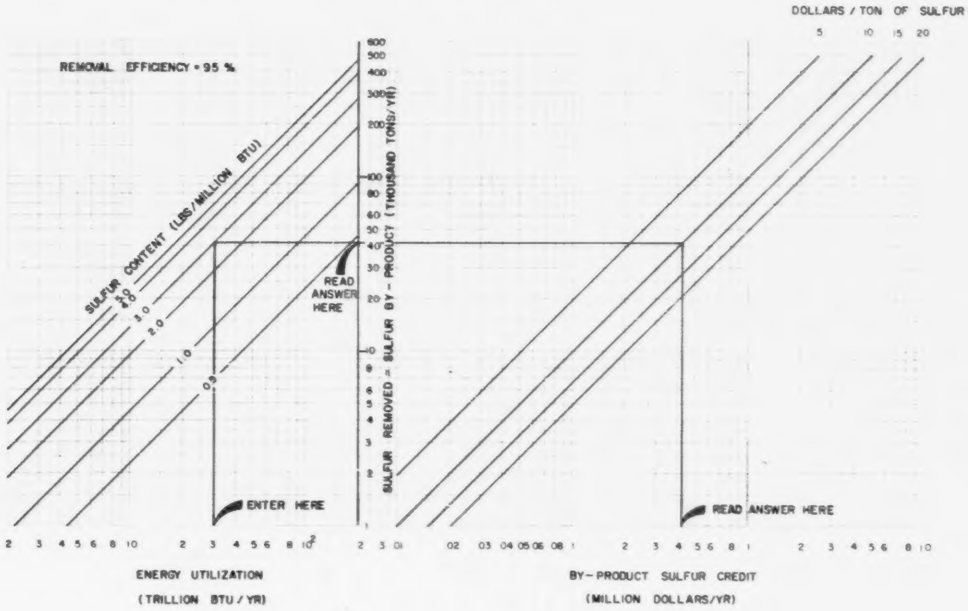


MOLTEN CARBONATE

B-OTHER OPERATING COSTS



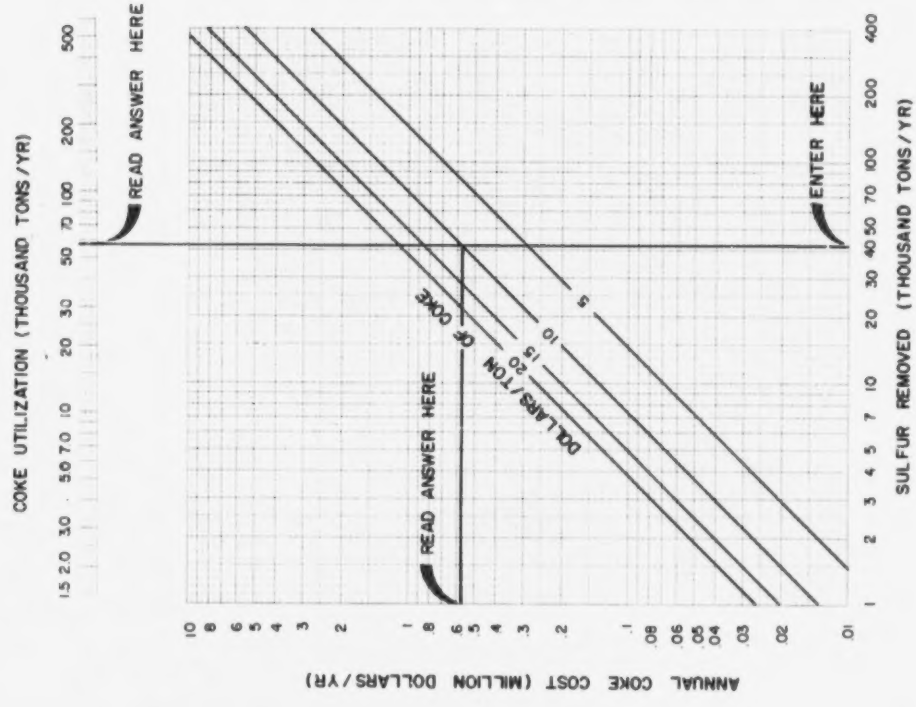
C-BY-PRODUCT SULFUR CREDIT



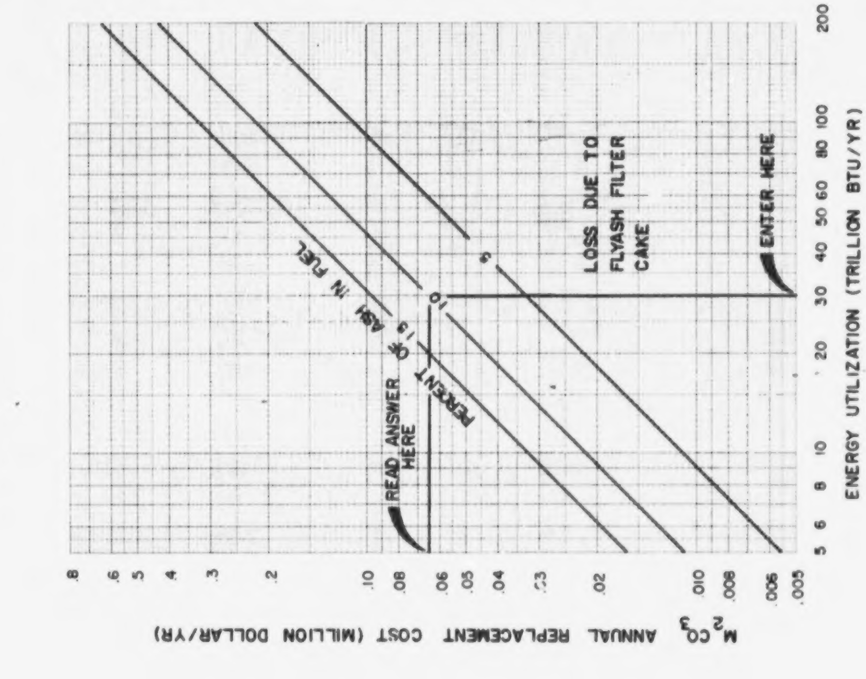
MOLTEN CARBONATE (CONTINUED)

D-RAW MATERIAL COSTS

I. COKE



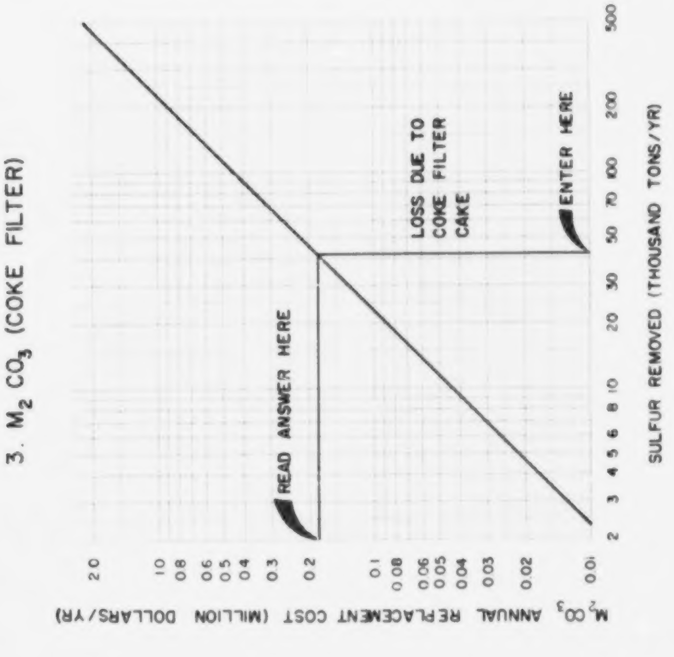
2. M_2CO_3 (FLYASH FILTER)



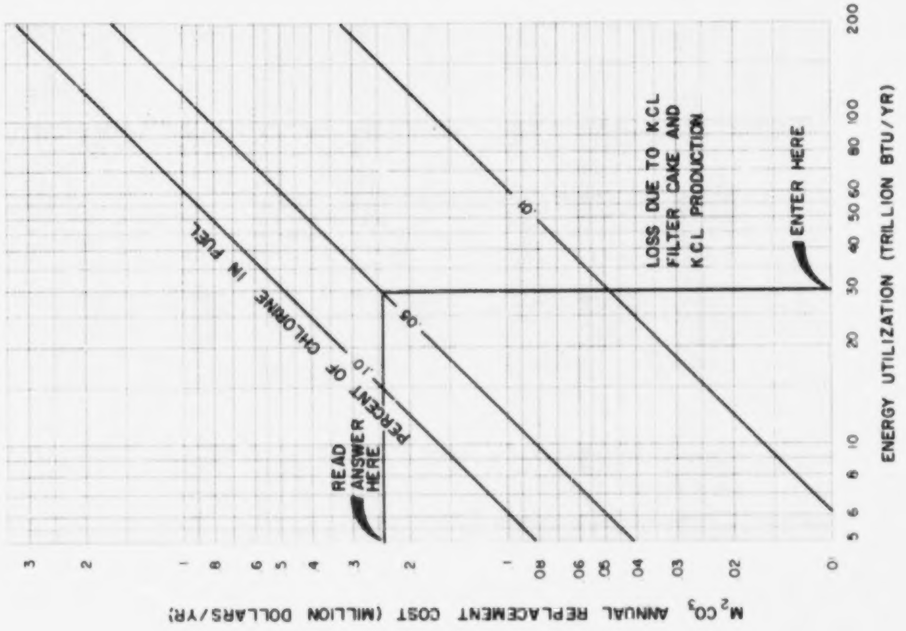
MOLTEN CARBONATE (CONTINUED)

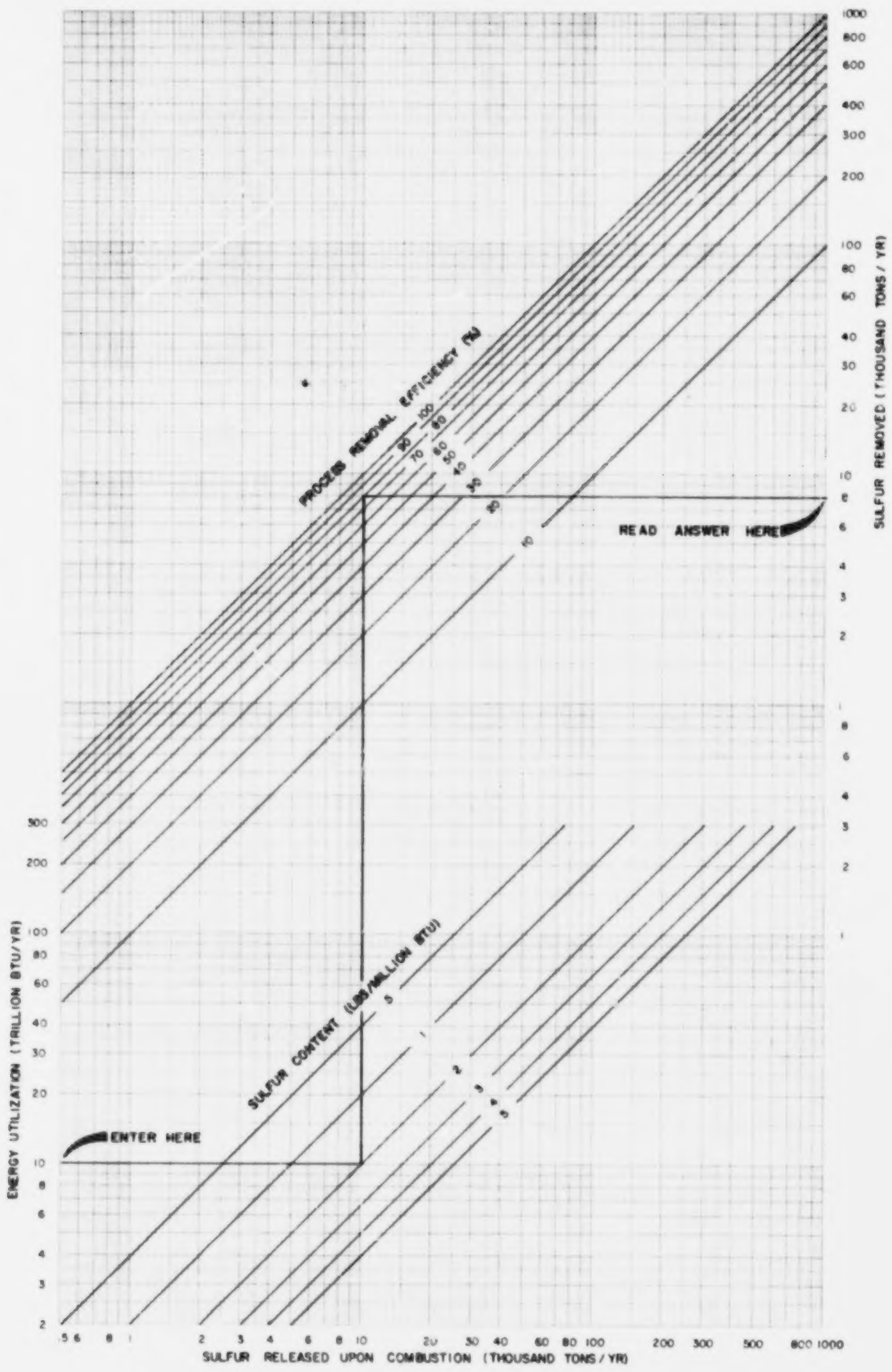
D - RAW MATERIAL COSTS (CONTINUED)

3. M_2CO_3 (COKE FILTER)

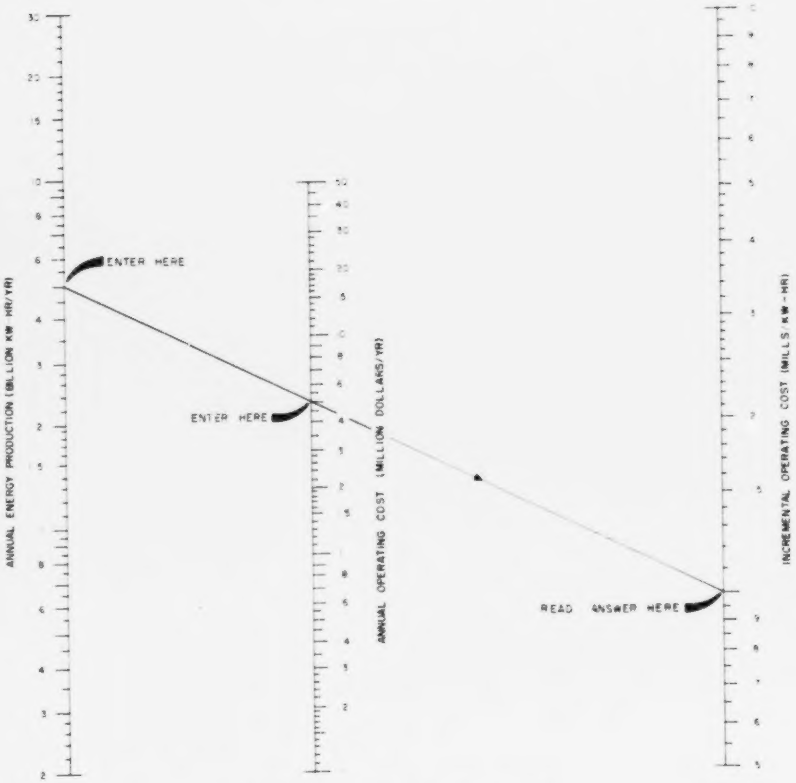


4. M_2CO_3 (KCL FILTER)





SULFUR REMOVED

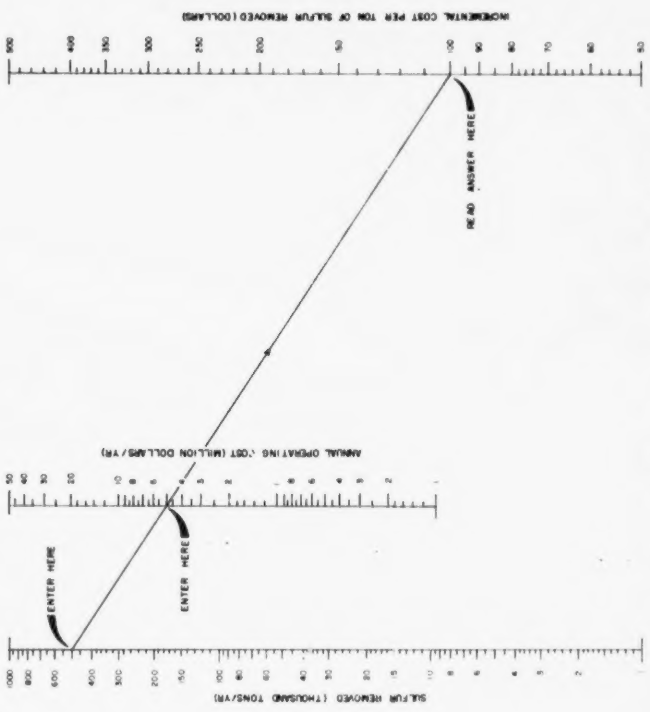


COST PER KILOWATT-HOUR

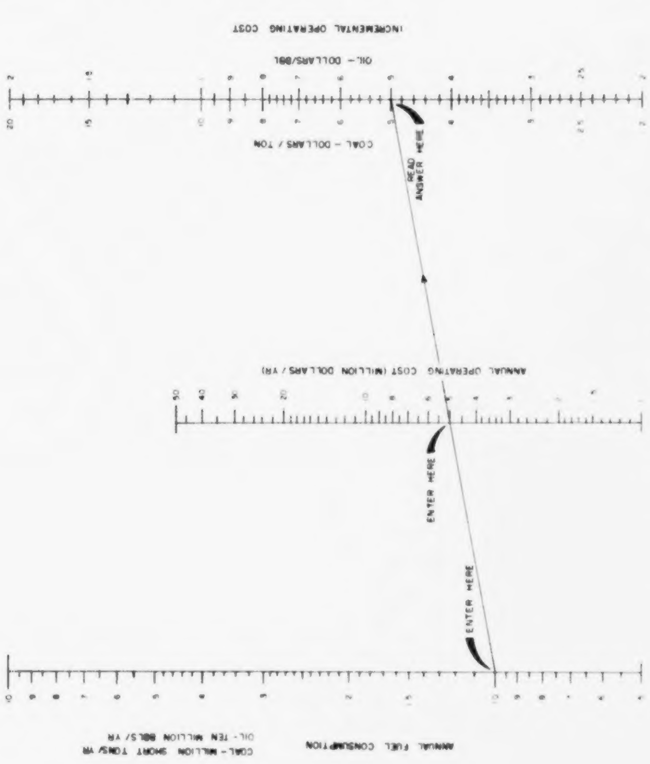
TYPICAL PROCESS REMOVAL EFFICIENCIES

(%)

DRY LIMESTONE	(VARIABLE SEE PAGE 5)
WET LIMESTONE	85.0
MAGNESIUM OXIDE	95.0
CATALYTIC OXIDATION	90.0
MODIFIED CHAMBER	87.5
MOLTEN CARBONATE	95.0


INCREMENTAL COST PER TON OF SULFUR REMOVED (DOLLARS)
ANNUAL OPERATING COST (MILLION DOLLARS/YR)
SULFUR REMOVED (THOUSAND TONS/YR)
ENTER HERE
ENTER HERE
READ ANSWER HERE

COST PER TON OF SULFUR REMOVED


INCREMENTAL OPERATING COST
OIL – DOLLARS/BBL
COAL – DOLLARS / TON
ANNUAL OPERATING COST (MILLION DOLLARS / YR)
ANNUAL FUEL CONSUMPTION
COAL – MILLION SHORT TONS/YR
OIL – TEN MILLION BBLS/YR
ENTER HERE
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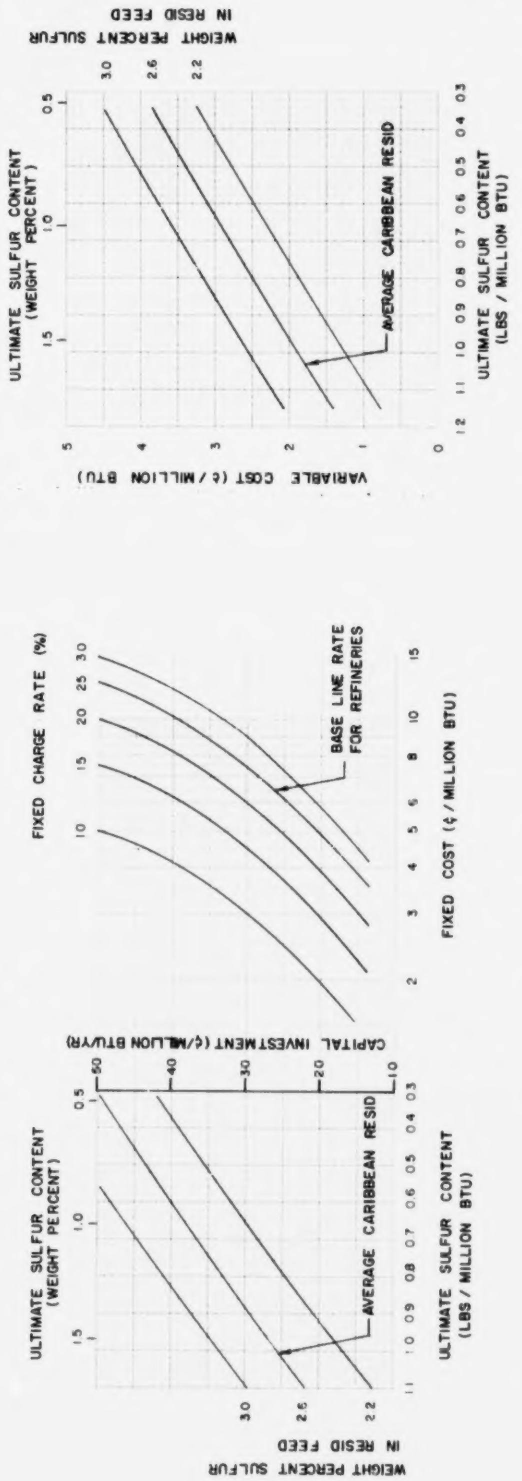
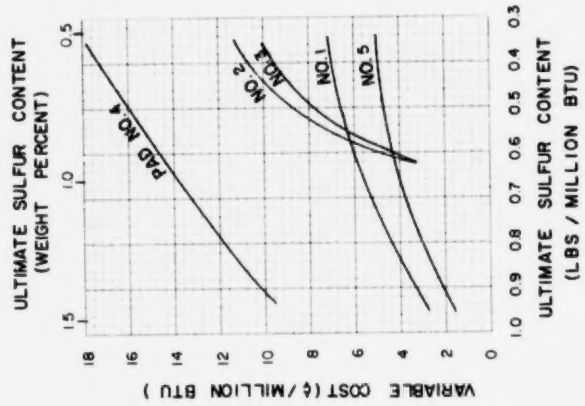
READ ANSWER HERE

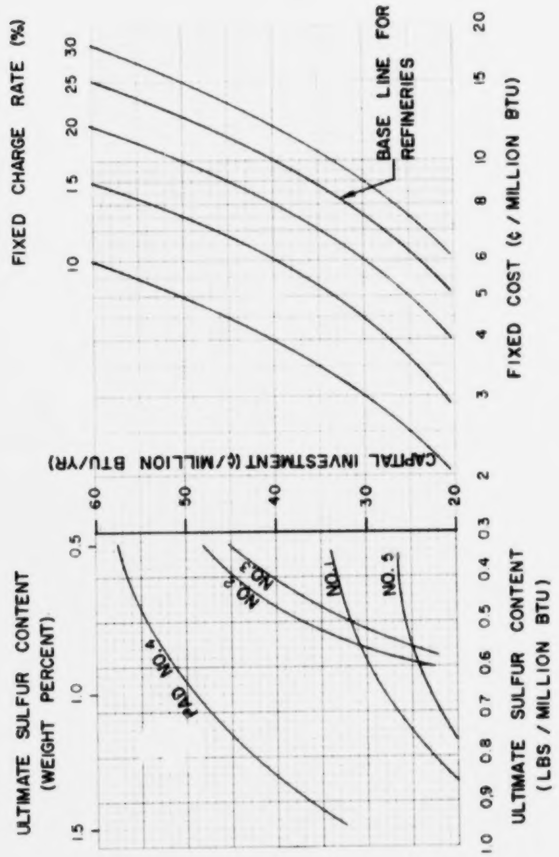
COST PER UNIT FUEL CONSUMPTION

CONVERSIONS (CONTINUED)

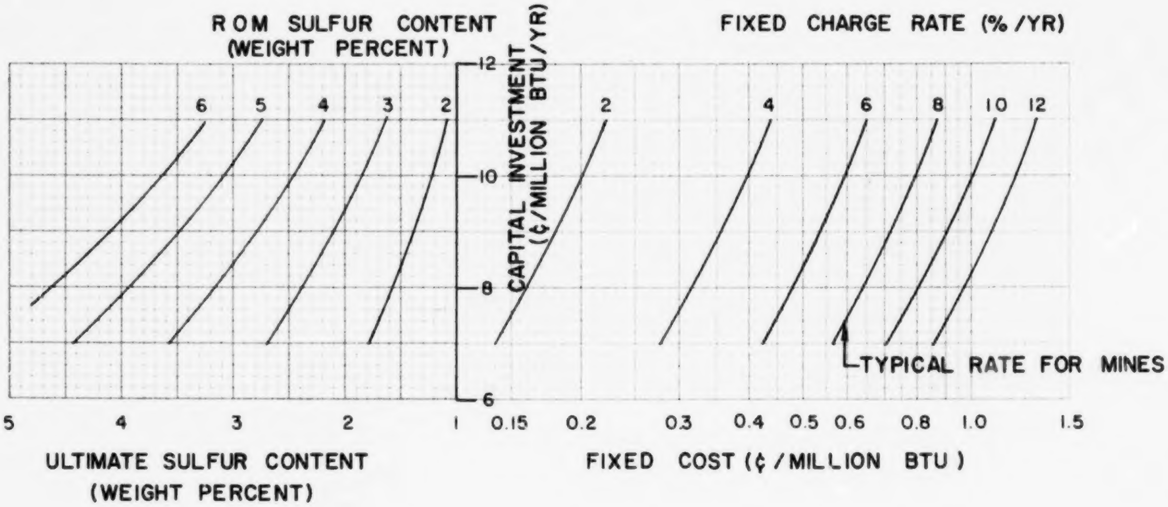


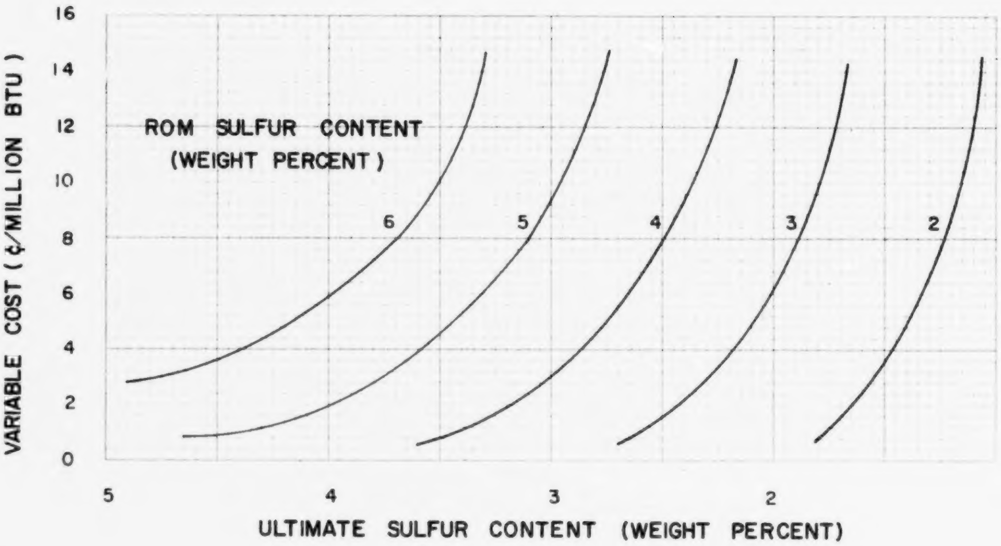
OIL DESULFURIZATION - CARIBBEAN REFINERY

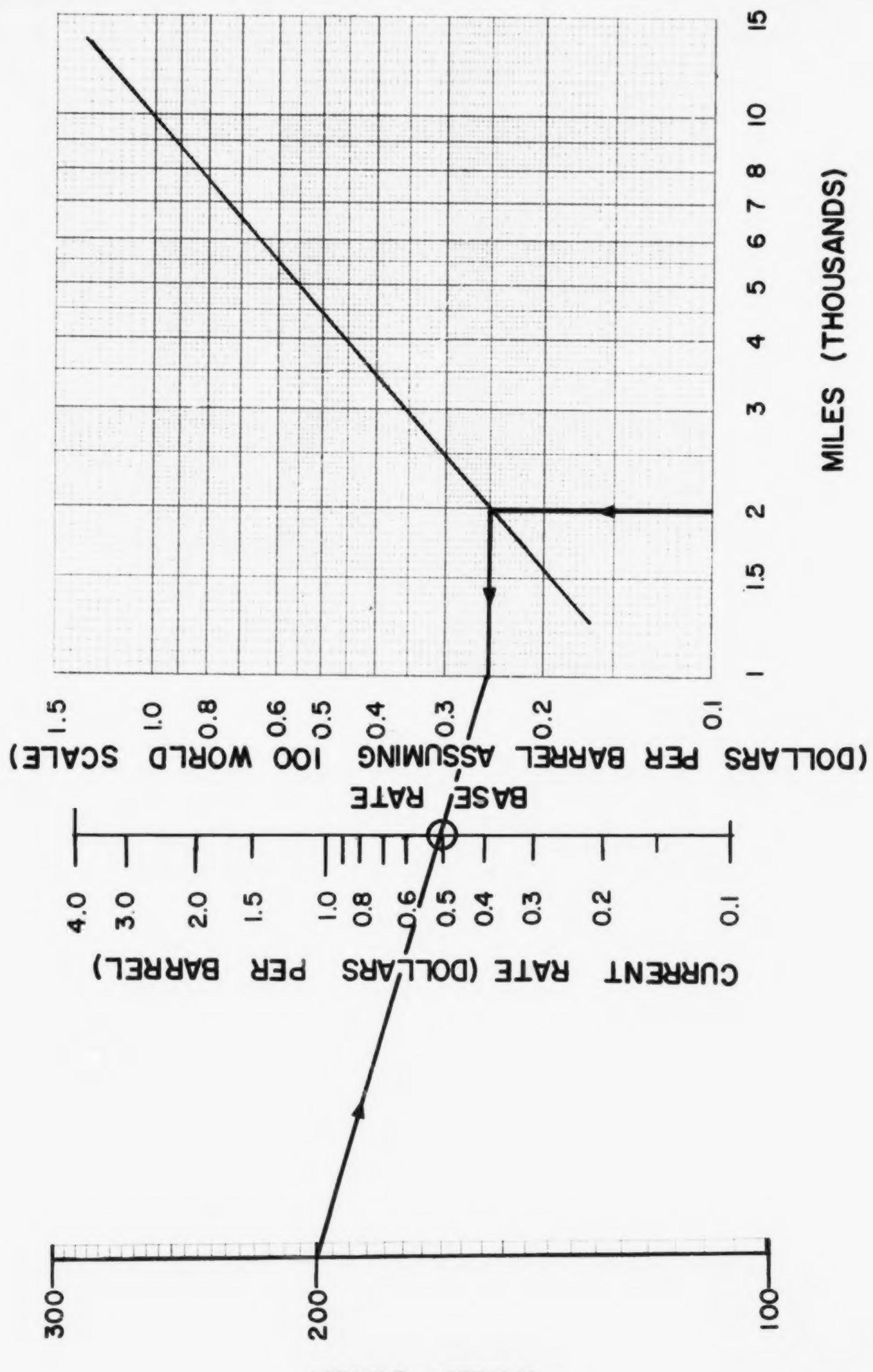




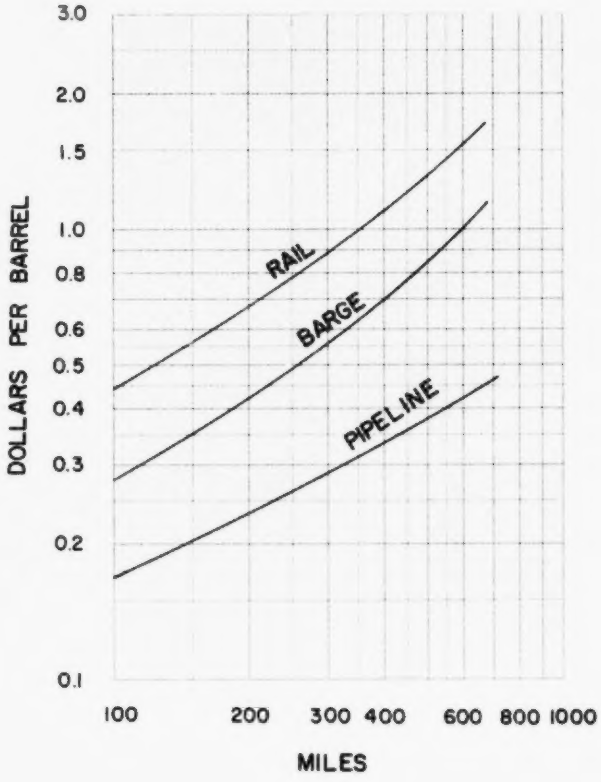
OIL DESULFURIZATION - DOMESTIC REFINERY


ROM SULFUR CONTENT
(WEIGHT PERCENT)
6
5
4
3
2
CAPITAL INVESTMENT
(¢/MILLION BTU/YR)
12
10
8
6
FIXED CHARGE RATE (% /YR)
2
4
6
8
10
12
TYPICAL RATE FOR MINES
5
4
3
2
1
ULTIMATE SULFUR CONTENT
(WEIGHT PERCENT)
0.15
0.2
0.3
0.4
0.5
0.6
0.8
1.0
1.5
FIXED COST (¢ / MILLION BTU)


VARIABLE COST (¢/MILLION BTU)
16
14
12
10
8
6
4
2
0
ROM SULFUR CONTENT
(WEIGHT PERCENT)
6
5
4
3
2
5
4
3
2
ULTIMATE SULFUR CONTENT (WEIGHT PERCENT)



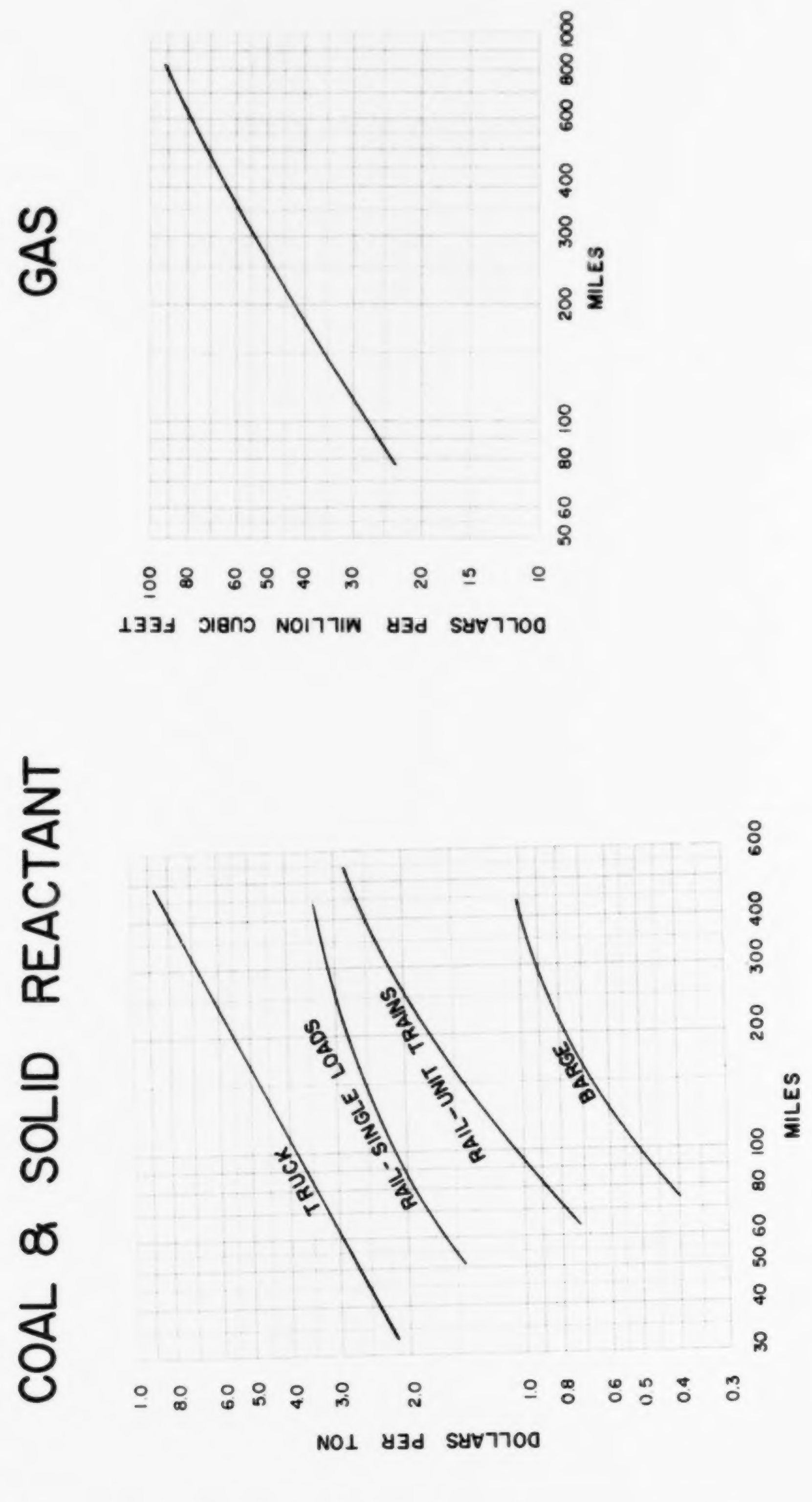
OIL TRANSPORT (FREIGHTER)



MILES (DAYS ROUND TRIP)
TO U.S.A., EAST COAST, FROM:

PERSIAN GULF (VIA CAPETOWN)	12,000 (65)
EAST MEDITERRANEAN	5,200 (30)
LIBYA	4,500 (28)
ALGERIA	3,800 (24)
CARIBBEAN	2,000 (13)

OIL TRANSPORT (OTHER MODES)



COAL AND GAS TRANSPORT

III. PARAMETRIC APPLICATIONS

One of the more revealing applications of the process nomograph system is in preparing parametric curves. Because of the ease of manipulating the basic economic variables, one can perform a parametric study in a relatively short period of time. Six such parametric studies were made to exemplify the kind of comparative analysis the nomographs make possible. The cost of SO_2 control in terms of mills/Kw-hr were calculated for over a dozen cases. Each case was run through each of the six process nomographs. The method involved varying one parameter at a time, keeping all others constant. The cost impact of the varying parameter was thus revealed. In Figure III-1, curves A-F summarize the parametric results. Each curve is based on the same basic set of parameters which are summarized below:

Base Case Inputs

Net generating capacity	500 Mw
Sulfur content	3.5 wt-%
Load factor	75%
Additive costs	Typical (see Table II-3)
By-product revenue	Typical (see Table II-3)
Fixed charge rate	20%/yr
Plant heat rate	11,000 Btu/lb (coal)

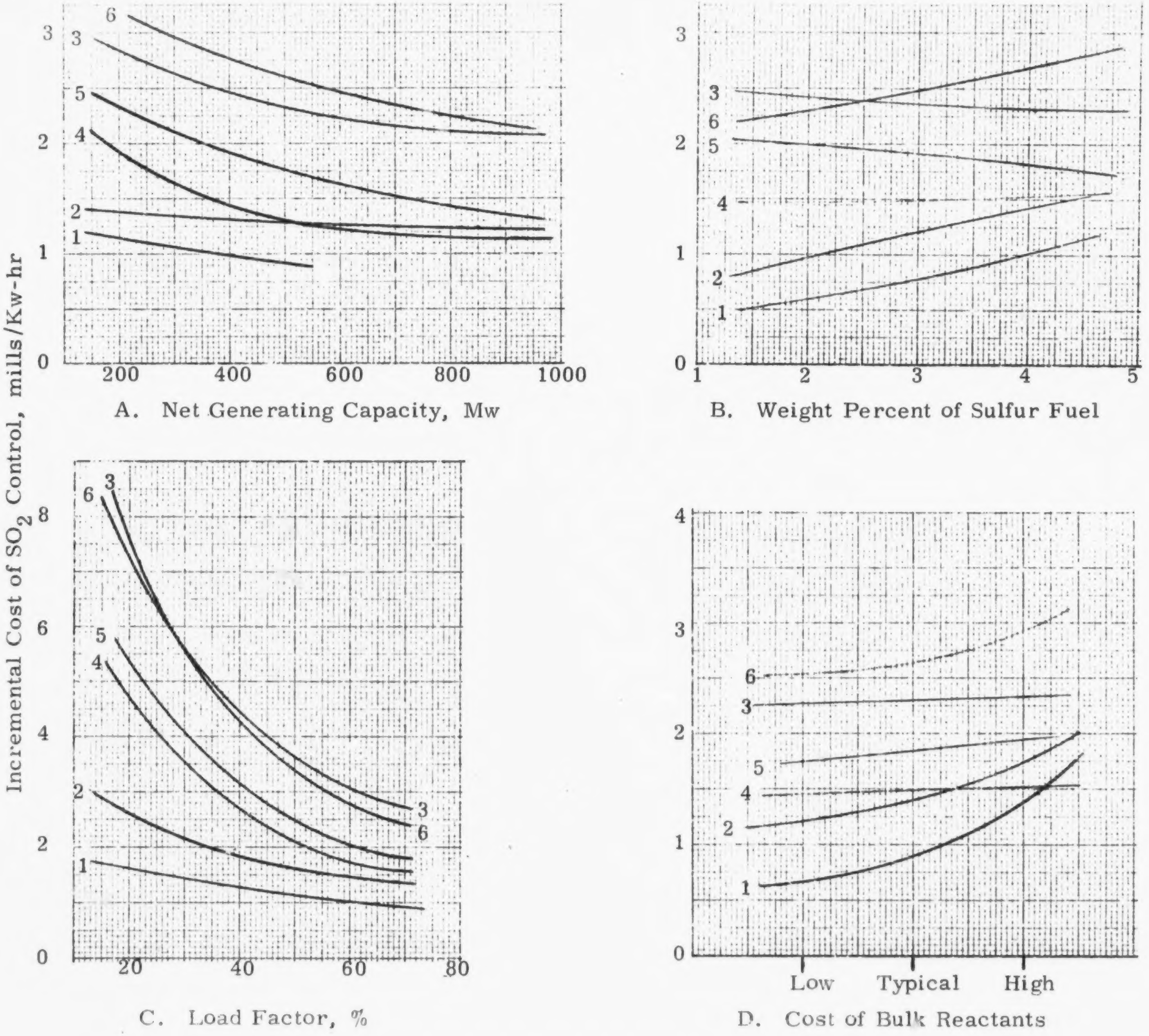
The base case parameters were varied as follows:

A.	Net generating capacity	200, 500, 1000 Mw
B.	Sulfur content	1, 3.5, 5 wt-%
C.	Load factor	20, 50, 75
D.	Additive costs	Low, typical, high (see Table II-3)
E.	By-product revenue	Low, typical, high (see Table II-3)
F.	Fixed charge rate	5, 10, 20 %/yr

These input variations were entered into each of the six process nomographs. The numbered curves appearing in Figure III-1 correspond to one of the six processes as follows:

1. Dry limestone
2. Wet limestone
3. Magnesium oxide
4. TYCO modified chamber
5. Catalytic oxide
6. Molten carbonate

In studying the resulting variations in control costs, many significant conclusions can be drawn. For instance, in Curve A, note the significant increase in cost of Processes 3-6 as plant size is decreased. Process 2, however, is less affected over the range. Note also that costs of Processes 2, 4, and 5 approach one another at the 1000 Mw level. When sulfur content was varied, costs were affected as shown in Curve B. Note that the limestone processes are affected most adversely. The other processes having some sulfur by-product are generally affected less. Note that Processes 3 and 5 actually decrease in cost as sulfur content is increased. Curve C shows the dramatic change in cost when load factor is varied. As is expected, the limestone processes are least affected since they are the least capital intensive processes. Similar behavior patterns are evident when the cost of bulk reactants, fixed charge rate, and the value of by-products are varied (Curves D-F). Thus, the nomographs provide a flexible and easy-to-use method of analyzing any of the important economic parameters.



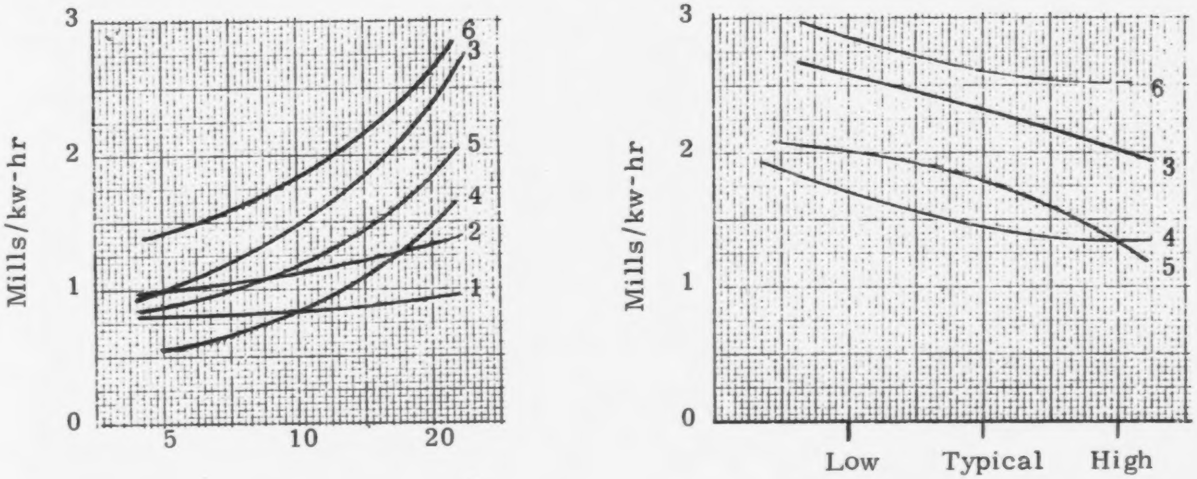
A. Net Generating Capacity, Mw

B. Weight Percent of Sulfur Fuel

C. Load Factor, %

D. Cost of Bulk Reactants

Figure III-1. Parametric Curves



E. Fixed Charge Rate, %/yr

F. By-Product Value

Figure III-1. Parametric Curves (Continued)

IV. DISCUSSION

A. Basic Methodology of Nomograph Preparation

The overall purpose of the nomographs is to present fixed capital and annual operating costs for six SO_2 removal processes. These costs are presented as a function of the major independent variables such as plant size, plant load factor, plant efficiency, and the weight-percent of sulfur in the fuel. Fixed capital and annual operating costs were based primarily upon the M.W. Kellogg Company unified cost estimates. These estimates were for plant capacities and coal sulfur levels as shown in Table IV-1. Other sources for costs were employed such as the reports from the process developers, EPA, TVA, and other consultants. Cost estimates were also performed by Hittman Associates to extend plant size from 100 to 2000 Mw and sulfur content from 0.5 to 5 lb $S/10^6$ Btu.

A cost sensitivity analysis was performed in which each independent parameter was ranked in terms of its significance to overall cost. If varying the parameter from its minimum to its maximum possible value significantly affected the total operating cost, it was introduced into the nomographs as a variable. If it was not significant, a mean value was selected and it was handled as a constant. For example, the variability of operating labor rates always resulted in only a small overall effect on operating cost and therefore was fixed at the average rate of $4.50/hr. This analysis was necessary to minimize the number of variables for practical application.

The fixed capital costs for each process are presented as a function of the volumetric flow rate of the flue gas into the system in cubic feet per minute (scfm) at standard conditions. Flue gas flow rate is determined as a function of plant capacity (Mw) and plant heat rate (Btu/Kw-hr) in auxiliary nomograph No. 1. In most of the processes, fixed capital costs also depend upon the sulfur content of the fuel. For coal, sulfur content is determined from the following equation:

$$SC = \frac{\text{weight-percent of sulfur x } 10^4}{\text{heating value of the coal}}, \text{ lb } S/10^6 \text{ Btu}$$

Sulfur content is plotted for coal and oil on nomograph page 4.

TABLE IV-1. NOMOGRAPH PARAMETERS BASED ON THE KELLOGG ASSUMPTIONS

Case	Plant Capacity Mw	Wt-% of Sulfur in Coal	Sulfur Content lb S/10^6 Btu	Energy Utilization 10^{12} Btu/yr	Flue Gas Flow Rate 10^6 scfm
1	175	3.5	2.92	11	.32
2	350	2.	1.67	22	.64
3	350	3.5	2.92	22	.64
4	350	5.	4.17	22	.64
5	400	3.5	2.92	25	.72
6	700	3.5	2.92	44	1.27
7	1000	2.	1.67	63	1.8
8	1000	3.5	2.92	63	1.8
9	1000	5.	4.17	63	1.8

NOTES:

Plant Heat Rate = 9000 Btu/kw-hr
Overall Plant Efficiency = 38%
Heating Value of Coal = 11,980 Btu/lb as received
Plant Load Factor = 80% (7000 hr/yr)
Annual Fuel Consumption for Base Case = 2.63 x 10^6 tons/yr
Cases 1-4, Dry Limestone Process only
Cases 5-9, All Other Processes
Case 8, Base Case

In the nomographs, the total annual operating cost is determined for a given condition by summing the fixed cost, the raw materials costs, and the other operating costs, and by subtracting the credit for by-products resulting from the process. The fixed cost in dollars per year depends upon the capital cost and the fixed charge rate. The fixed charge rate in percent per year is a function of the interest rate, amortization period, and the recurring cost percent. It is determined by adding the recurring cost percentage to the capital recovery factor obtained from nomograph page 1. Other operating costs include utilities, direct labor, supervision, maintenance, overheads, and waste disposal. Other operating cost magnitudes also account for other credits that are unique for each process such as stack gas cooling and boiler feedwater heating.

The annual operating costs and credits in the process nomographs depend upon plant capacity through the parameter, energy utilization. Energy utilization is equal to the product of the annual fuel consumption (at a given plant load factor) and the heating value of the coal. It is determined from nomograph page 2. These costs also depend upon the parameters: sulfur content of fuel, ash content of the fuel, chloride content of the fuel, NO_x content of the flue gas, raw material unit costs, by-product unit costs, etc.

In addition to the Plant Size and Fuel Consumption Nomograph (pages 1-4) and the six SO_2 removal process nomographs, two other nomographs were constructed. On pages 27-29, operating costs can be converted to other parameters for comparison purposes. The sulfur emitted from the stack is also determined as a function of sulfur content, energy utilization, and SO_2 removal efficiency. On pages 30-35, the costs of oil and coal desulfurization are computed. Also, the costs of oil, coal, and gas transportation are determined.

B. Basis of Process Cost Estimates

The basic approach to estimating cost involved the comparison between a new plant with and without SO_2 control. Conventional equipment which is eliminated due to the addition of the control equipment was credited to the process capital cost. The increase or decrease in the cost of dust collectors, induced draft fans, ductwork, insulation, etc., required when adding the control equipment was considered a debit or credit to the control process.

Cost estimates have been performed by EPA, Hittman Associates, M. W. Kellogg Company, TVA, and Singmaster & Breyer, along with the process developers. In general, these estimates were difficult to compare since no single set of cost estimating techniques were used. It was found that comparisons between estimates could be very misleading. A large amount of effort was sometimes necessary in order to discover the reasons for various discrepancies in the data. These difficulties point to the need for a uniform cost guideline. Such a guideline would allow comparisons between estimates, would allow an individual to adjust an estimate based on his own needs and would eliminate many potential errors and omissions.

Primarily because estimates from different organizations were difficult or impossible to compare on a single basis, it was decided that the most uniform set of estimates would be utilized as a basis for the nomographs. It was reasoned that the process-to-process comparisons would only be fair if a single methodology was used. In reviewing all the estimating work completed in 1971, it was found that the M. W. Kellogg Company had provided EPA with estimates of the six stack gas processes (among others) based on uniform methods for estimation and presentation. These estimates were for a base case of 1000 Mw plant size and a 3.5 wt-% sulfur content in the fuel. These estimates were used as the basis of the parametric analysis performed by Hittman Associates.

The breakdown of the capital cost estimates performed by the Kellogg Company for the base case is given in Table IV-2. The base case for the dry limestone process was a plant capacity of 350 Mw and a coal whose sulfur content was 3.5 percent by weight. The base case for the remaining five SO_2 removal processes was a 1000 Mw plant burning the same coal. The costs of induced draft fans and dust collection equipment required with the process are included in the major equipment cost. Credit for the cost of these components sized for a plant with no controls are given at the bottom of the table. A contractor's fee was included under "Other Costs." The fixed capital cost or net cost investment does not include any contingency or interest during construction.

The annual operating costs for the base case of each process are presented in Tables B-1 through B-6 of Appendix B. These estimates and the fixed costs of Table IV-2 were transmitted by the Kellogg Company to Hittman Associates in Reference 1. The operating cost assumptions are indicated in

TABLE IV-2. SO_2 REMOVAL PROCESSES - BASE CASE (Costs in $\$10^3$)

	Molten Carbonate	Wet Limestone	Cat-Ox New or Retrofit	Tyco	Magnesia Case	Dry Limestone 350 Mw
Major Equipment Cost:						
Direct materials	3,637	5,395	13,525[c]	9,754	8,422	664
Subcontracts	4,231	324	1,360	6,949[d]	1,665	91
Ductwork, Duct Valves, and Expansion Joints	3,565[a]	937	4,231[a]	811	1,182	124
Piping, Electrical, and Instruments	2,830	1,323	1,940	2,360	2,173	166
Foundations, Concrete and Steel Structures, Buildings, Paint, and Insulation:						
Direct materials	474	340	440	410	470	76
Subcontracts	673	170	340	290	290	70
Subtotal - Materials and Subcontracts	15,410	8,489	21,836	20,574	14,202	1,191
Other Costs:						
Construction forces, home office engineering, procurement, insurance, startup, and contractor's fee	14,490	6,711	17,064	10,226	11,398	1,409
Subtotal Investment Cost for Process Plant	29,900	15,200	38,900	30,800	25,600	2,600
Other Process Units Investment Costs:						
Claus sulfur plant	1,600					
Sulfuric acid plant including H_2SO_4 storage			1,400[b]	1,300[b]	5,600	
Electrostatic precipitator	7,500[f]		7,500[f]			2,400[e]
Deduct Standard Power Plant Induced Draft Fans and Ducts	(2,300)	(2,300)	(2,300)	(2,300)	(2,300)	(700)
Deduct Standard Power Plant Electrostatic Precipitator[e]	(2,100)	(2,100)	(2,100)	0	(2,100)	(900)
NET COST INVESTMENT	34,600	10,800	43,400	29,800	26,800	3,400

NOTES:

( ) = Deduction

[a] Costs also include duct insulation.

[b] Cost is for sulfuric acid storage only.

[c] Cost includes V_2O_5 catalyst.

[d] Cost includes nitric acid storage.

[e] 300°F electrostatic precipitator.

[f] 900°F electrostatic precipitator.

the tables. Labor rates were assumed based on a plant location in Cincinnati, Ohio. The stream time corresponds to a plant load factor of 80 percent. The percentage of fixed capital investment accounting for maintenance labor and materials was a variable. The unit cost for electric power under cost of utilities was assumed to be $0.00675/Kw-hr. Solid waste disposal was assumed to cost $0.50/ton. The fixed charge rate was assumed to be constant at 18.22 percent per year in these estimates. This parameter can be varied in the nomographs when the fixed cost is required under various economic conditions. The total operating cost is defined as being the net production cost wherein account has been taken of the various credits.

Cost estimate data points were extended beyond the base case through an additional effort performed by M. W. Kellogg Company. Reference 2 presents the effects on costs of varying the plant capacity and the sulfur content of coal. Results of the parametric study are shown for each SO_2 removal process in Tables B-7 through B-12 of Appendix B. One important assumption in the study was that the concentration of sulfur dioxide emitted in the power plant effluent gas be held constant for all cases at approximately 250 ppm. In the tables, the percent of sulfur removed from the flue gas or the SO_2 removal efficiency of each case differs somewhat, though not significantly, from that of the base case.

The total gross production cost less the credits yields the total net production cost for a given case. In determining credits, the unit costs for acid are for the concentrated acid. This magnitude is reduced proportionately to obtain the unit cost at the lesser concentration.

C. Discussion of Processes

1. Dry Limestone Process

a. Process History. The first active program in the United States for development of a dry limestone injection process was initiated in 1964 by the Process Control Engineering Program of EPA. As part of this program, a conceptual design study was conducted by the TVA for the development of a large scale prototype. This prototype design was scheduled for testing in 1969 at TVA's Shawnee and Paradise plants in Kentucky.

Presently, the process is undergoing tests at the Shawnee plant near Paducah, Kentucky (Ref. 3). Current demonstrations are achieving from 20 to 30 percent control of sulfur oxides. The EPA has indicated that an ultimate goal of 50 percent efficiency is being sought. The current program is for \$3.3 million and 18 months (Ref. 4).

The process can use dolomite, which is a mineral containing equal parts of calcium and magnesium carbonate (commonly symbolized by $MgCO_3 \cdot CaCO_3$), or any alkaline earth such as limestone which, by definition, contains more than 50 percent calcium carbonate ($CaCO_3$). Limestones are abundant in most areas of the country, but their compositions vary widely. Dolomite limestones are those containing an appreciable percentage of magnesium carbonate ($MgCO_3$).

b. Process Description. The process is initiated with the limestone and/or dolomite being pulverized and fed into the high-temperature combustion zone of the furnace where it is calcined (oxidized) to the active oxide forms CaO and MgO. Figure IV-1 provides a schematic of the process. The reactions are as follows:

$$CaCO_3(s) \longrightarrow CaO(s) + CO_2(g)$$

$$MgCO_3(s) \longrightarrow MgO(s) + CO_2(g)$$

The reaction of the pulverized-calcined additives with SO_2 and oxygen at temperatures above 1200°F forms gypsum or calcium sulfate ($CaSO_4$) and magnesium sulfate ($MgSO_4$) which, along with the unreacted limestone and/or dolomite and the normal particulates (e.g., fly ash), are carried out of the furnace in the flue gases. The calcium and/or magnesium sulfates are removed with the fly ash in conventional dry scrubbers, either electostatic precipitators or cyclone separators. Additional electrostatic precipitator capacity may be required to maintain the particulate below regulated values. The gypsum and other sulfates, along with the unreacted limestone, are handled as waste along with the fly ash. Additional disposal capacity will be required to handle the increased waste. Because of the chemical composition of the resultant wastes, a water pollution problem may exist, depending upon the disposal system design and local conditions. This is especially true in the case of a sluice/settling pond system in which pond overflows due to heavy rains may damage stream ecology. This problem must be handled on a site-by-site basis.

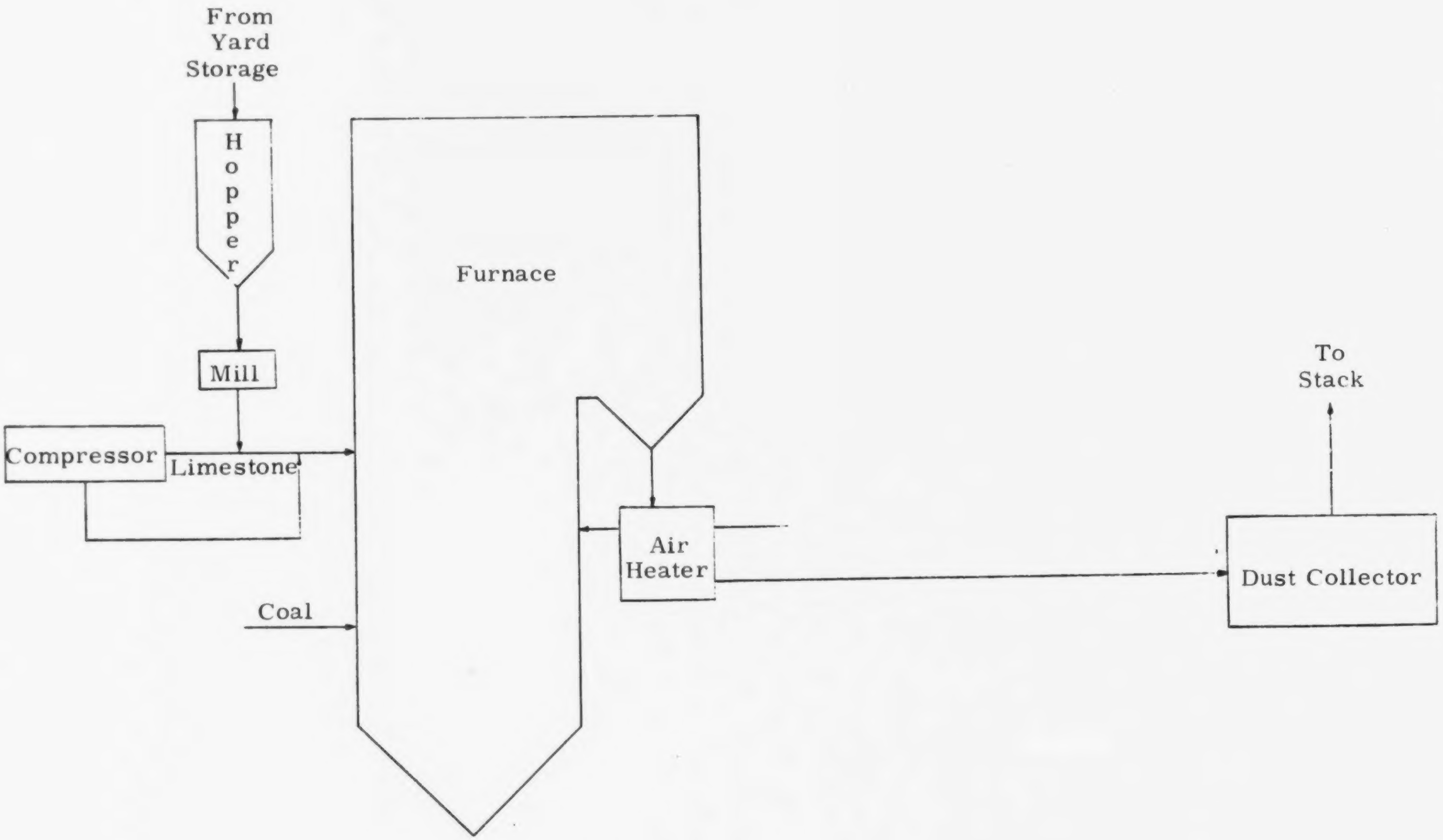

From Yard Storage
Hopper
Mill
Compressor
Limestone
Coal
Furnace
Air Heater
Dust Collector
To Stack

Figure IV-1. Dry Limestone Injection Process

c. Process Economics. Of all the processes, the dry limestone process is the least complex. It requires a minimum amount of equipment and is, therefore, the least capital intensive. It is also the process most easily retrofitted on an existing plant. External receiving hopper, storage facilities, pulverizer, and injection system plus conveyors between these units are all that are necessary for limestone injection. A more efficient dust collector and higher capacity disposal system make up the balance of the components.

The capital and variable costs of the system are displayed on nomograph pages 5-9. Because the selection of the stoichiometric ratio (1.0 stoichiometric refers to the amount of reactant required to eliminate all of the SO_2 assuming complete carbonate utilization) is known to significantly affect both SO_2 removal efficiency and operating cost, it has been introduced as a variable. Generally, a range of between 1.0 and 3.0 stoichiometry is selected. The selection of values higher than 3.0 normally results in an unacceptable amount of limestone entering the system. This material may overwhelm the dust collection equipment and/or result in excessive corrosion or erosion of boiler internals.

Graph A, which shows process efficiency versus stoichiometric ratio, is a reasonable interpretation of currently available date (Refs. 5 through 8). Efficiency varies widely because of the many contributing factors. Limestone chemical and physical characteristics cause significant variations in sorption efficiency. Smaller limestone particle size results in higher sorption efficiency. The time-temperature profile of the injected limestone particles has a significant effect on the calcination/sorption processes. The distribution of the particles as they enter the fire box also contributes to the variability in SO_2 removal efficiency. These factors and others are being studied in the attempt to increase the process effectiveness.

Graph B calculates the limestone use factor (LUF). The stoichiometry, the amount of sulfur in the fuel, and the limestone purity determine the amount of limestone used per Btu of energy liberation. The factor is defined below:

$$LUF = \frac{S \times R \times CSR}{P}$$

where:

LUF = limestone use factor, lb of limestone/10^6 Btu

S = sulfur content in fuel, lb of sulfur/10^6 Btu

R = stoichiometric ratio

CSR = ratio of $CaCO_3$ to sulfur at 1.0 stoichiometry, 3.12

P = limestone purity ≈ percent of $CaCO_3$ in limestone, %/100

The LUF is utilized a number of times in obtaining cost data. Because of the similarities in variables between the dry and wet limestone processes, the LUF graph is used for both processes.

Graph C allows calculations of the capital and fixed costs. The major variables affecting capital cost are the limestone use factor and plant flow rate. Graph E allows for the adjustment of costs when the use of a limestone particle size smaller than normal is found to be desirable. The graph should be used in conjunction with test data which shows a definite improvement in efficiency with finer grinding. If no test data exist, it is best to overlook the correction and assume standard grinding (80 percent of the material ground to less than 44 microns, which is equivalent to passing a No. 325 screen).

Graph F accounts for limestone usage. The nomograph calculates the amount of limestone used per year and the annual additive cost. The price per ton of limestone includes the following:

(1) FOB limestone cost
(2) Transportation cost
(3) Incremental grinding cost, if any
(4) Cost of disposal

The capital and operating costs for the disposal system are not included in the nomograph costs as separate items. These costs were handled on a cost-per-ton of limestone used. The Kellogg Company uniform estimate method used a cost of $0.50/ton for all processes which required the disposal of bulk waste. In particularly difficult locales which require more costly methods of disposal, the cost per ton may go as high as $1.00 to $1.50/ton. Since limestone cost calculations are also required for the wet limestone process, the graph is positioned to be utilized for both limestone processes.

All other operating costs other than the cost of limestone and the incremental grinding costs are included in a single curve (Graph D). The major portion of these other costs (which include labor, utilities, plant overhead, maintenance, etc.) is dependent on the quantity of limestone used. Thus, the parameters used to scale "other operating" costs are energy utilization and the LUF.

The total operating cost is found by adding the output of graphs C, D, E, and F.

2. Wet Limestone Process

a. Process History. The limestone-injection/wet-scrubbing process for SO_2 control was first researched in the United States by Wisconsin Electric Company and Universal Oil Products Company in 1963 and 1964.

In August 1967, Combustion Engineering received a contract from the Union Electric Company,on the strength of large-scale tests run at the St. Clair station of Detroit Edison Company, to install a full-scale combination wet-dry dolomite process on the 140 Mw unit No. 2 at the company's Meramec Station in St. Louis, Missouri. The unit has been operating only intermittently because of mechanical problems. Combustion Engineering installed another system at Kansas Power and Light's 125 Mw unit No. 4 at Lawrence, Kansas. This unit achieved continuous operation in 1970 (Ref. 9).

Both the Meramac and Lawrence systems have recently been down for modifications based on previous tests. Both are now in operation. In addition, Kansas Power and Light is just completing its 420 Mw Lawrence Unit No. 5 which is equipped with the same SO_2 removal system.

Bechtel Corporation, under contract to TVA and APCO, is currently installing a major test facility at a 175 Mw unit at Shawnee near Paducah, Kentucky. This facility, which is also capable of testing dry limestone systems, is being designed to test almost unlimited variations in equipment. Different wet liquor circuits, wet scrubbers, and limestone preparation subsystems can be tested in various combinations.

b. Process Description (Refs. 10-13). The Combustion Engineering process, which employs furnace injection of an alkaline earth followed by wet scrubbing,is shown in Figure IV-2. The pulverized additive calcines in the furnace and reacts with the combustion gases to form compounds of calcium and manganese. This removes 10 to 20 percent of the sulfur oxides, including most of the SO_3.

Flue gas containing unreacted SO_2 and calcined additive then passes through one or several air heat exchangers and into the wet scrubber. The scrubber may be of several designs. The more promising are the venturi, mobile-packing, and the flooded-bed scrubbers. Because particulate removal is also a requirement, these scrubber systems are being evaluated not only for their SO_2 removal effectiveness, but also for their ability in removing particulates.

In the scrubber, the calcined additive that had not combined with SO_2 in the furnace reacts with the water to form a slurry of reactive milk-of-lime (a suspension of calcium hydroxide in water). The process is

$$CaO(S) + H_2O \longrightarrow Ca(OH)_2(S) + 16\ Kcal$$

It might be noted that this process is accompanied by a threefold expansion in volume.

The $Ca(OH)_2$ (and $Mg(OH)_2$) reacts with the remaining SO_2 to form sulfates and sulfites of calcium and magnesium. Overall SO_2 removal efficiency of about 85 percent is typical. At the same time, water entrainment of fly ash removes particulate matter. Additive and slurry flow rates control the scrubber reaction.

The solution containing the reacted materials drains out the bottom of the scrubber to a tank (clarifier) or pond where the particulates settle. The clarified slurry is then available for recirculation.

For improved utilization of additive, provisions can be made for direct recycling of a portion of the scrubber effluent to above-bed sprays. The cleansed flue gas passes through a mist eliminator for removal of its remaining water and is then reheated. Reheat is necessary to protect upstream equipment such as induced draft fans and also to obtain better air dispersion of the effluent.

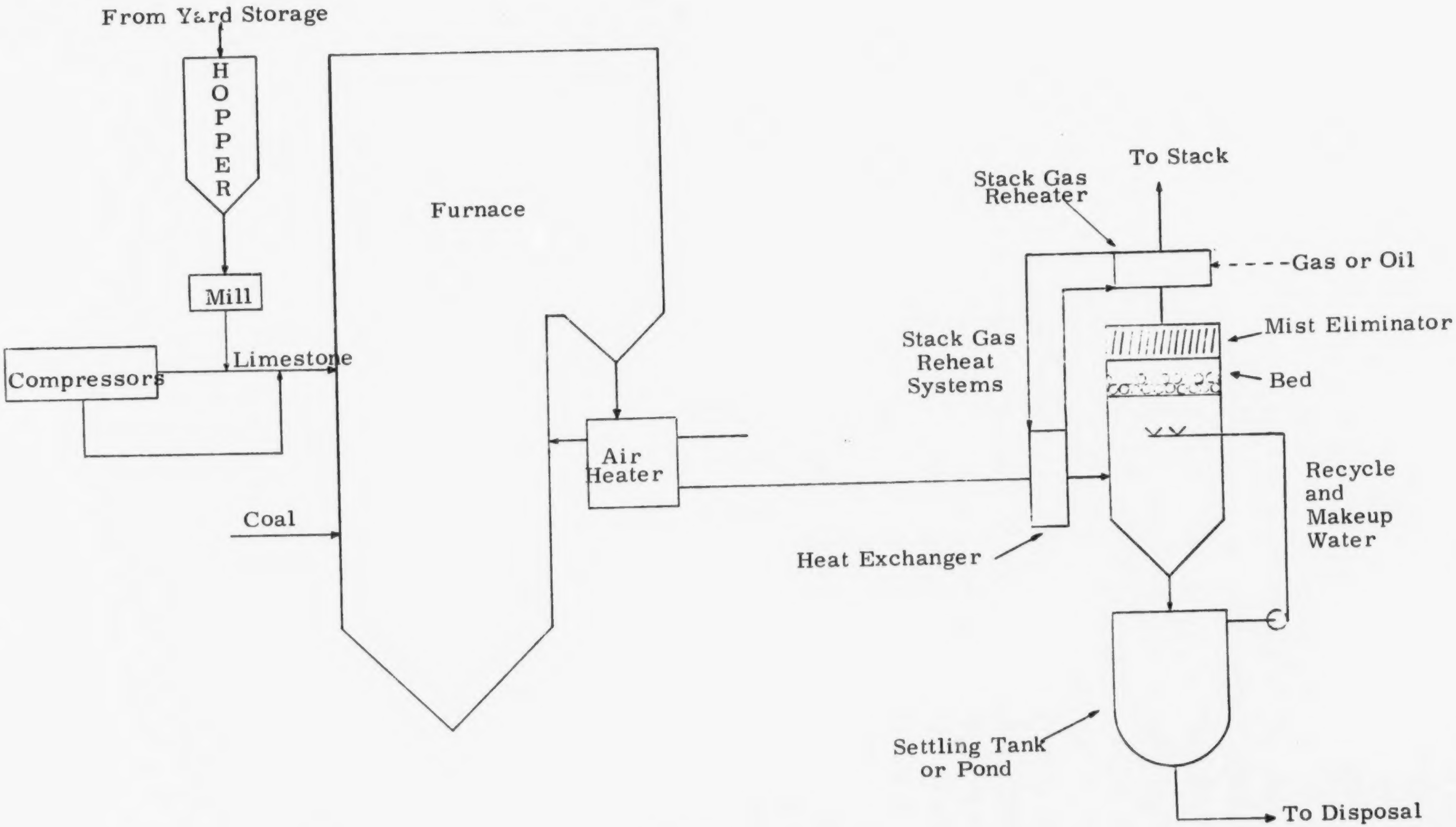

From Yard Storage
HOPPER
Mill
Compressors
Limestone
Coal
Furnace
Air Heater
Heat Exchanger
Stack Gas Reheat Systems
Stack Gas Reheater
To Stack
Gas or Oil
Mist Eliminator
Bed
Recycle and Makeup Water
Settling Tank or Pond
To Disposal

Figure IV-2. Wet Limestone Injection Process

Combustion Engineering operations for 1969 at Meramec and Lawrence, each cleaning 300,000 to 400,000 cfm of gas, revealed information and problems not evident during pilot plant operation. As a result, these systems have been modified in the areas of additive injection, gas distribution, and water control.

Operations when firing 3.4 percent sulfur coal show an SO_2 emission rate equivalent to burning coal with less than 0.5 percent sulfur (~85 percent efficiency). Greater than 99 percent of the particulate matter can be removed and nitrogen oxide emissions are cut by as much as 30 percent.

c. Process Economics. The wet limestone system requires relatively few major pieces of equipment. Its capital cost per kilowatt is intermediate between the dry limestone system and the more sophisticated by-product systems. The major capital cost element is the flue gas scrubber/demister component which amounts to nearly half of the equipment costs. The flue gas flow rate is the major determinant for sizing the scrubber. The next most costly items include the disposal/settling pond and associated piping, pumps, and controls. The costs of these components are dependent upon the distance the pond is placed from the power plant, the geology at the pond site, the sulfur content in the fuel, the stoichiometric ratio, and energy utilization of the plant. The capital costs of these components are not included in the nomographs since process wastes were accounted for by charging the process on a cents/ton of waste basis. Thus, the resultant capital cost estimate is 15 to 30 percent lower than the estimate which includes settling pond costs. The limestone storage, handling, and processing equipment make up the last major equipment items. These components typically amount to about 15 percent of the capital investment.

The nomograph was prepared using four expense categories including fixed cost, reheat credit, additive cost, and other costs. The fixed cost is the annual proration for all capital equipment and installation costs excluding settling pond costs. The additional labor and materials required for a retrofit installation were not factored into the fixed costs. If an existing plant was fitted with a process, the fixed cost may increase by 10 to 20 percent depending upon the modification required to install the process equipment.

The reheat system may either be a gas-gas or gas-liquid-gas heat exchanger system or it may use a direct natural gas heater. The estimate used for nomograph preparation includes the capital and operating cost for a gas-liquid-gas exchanger sized to obtain a 250°F stack temperature downstream of the process equipment. This is the maximum temperature required to assure reasonable plume dispersion characteristics. Circumstances may allow a reheat temperature below 250°F. A credit would have to be applied to account for the reduction in cost of the reheat system. The size of the credit depends upon flow rate and reheat temperature. The incremental fixed cost (assuming 15 percent fixed charge rate) and variable costs were estimated for various exchanger capacities and reheat temperature resulting in a credit nomograph which can be utilized if a reheat temperature of less than 250°F is used.

The additive cost is determined by the same method used in the dry limestone process. Because the wet limestone method involves higher fixed charges and has a lower limestone use factor, the significance of the additive costs are less significant. At the 500 Mw plant size, the additive cost will typically run about one-third of the total operating cost.

Other operating costs include labor, plant supplies, utilities, and indirect costs involved in operating the process. As in the dry limestone process, the major variables are energy utilization and the limestone use factor. Other costs typically amount to about one-third of the operating cost of a 500 Mw plant as shown in the example presented previously. Nomograph 3 calculates the above costs. Total operating cost is found by adding the output of graphs C, D, and F and subtracting the output of graph E.

3. Magnesium Oxide Process

a. Process History. A sulfur dioxide abatement process developed by the Chemical Construction Corporation (Chemico) utilizes magnesium oxide in the main reaction. Sulfur dioxide is a by-product which can be converted to sulfuric acid. Extensive pilot plant work has indicated the process to be technically and economically feasible. A prototype commercial system will be built at the Mystic No. 6 Unit of Boston Edison Company to handle flue gases from a 155 Mw boiler. Target date for completion and startup is approximately at the end of 1971, followed by a testing period of 24 months. The nearby sulfuric acid facility of Essex Chemical Company will be utilized to produce 98 percent concentrated sulfuric acid and to regenerate magnesium oxide.

The magnesium oxide process is also being developed by the Babcock and Wilcox Company. They built a wet scrubbing pilot plant consisting of three scrubbers. These three scrubbers consisted of a venturi-type particulate scrubber, a venturi-type absorber, and a tray-type absorber (floating bed absorber). Many short-term and extended tests were performed to determine the most satisfactory operating conditions for each scrubber.

b. Process Description. In the process, the flue gas can be first passed through a venturi-type particulate scrubber to remove greater than 99 percent of the fly ash. At the Mystic site, however, no prescrubber is employed since the boiler is oil-fired resulting in little fly ash. In Figure IV-3, flue gas containing the sulfur dioxide enters a venturi-type absorber in which it contacts the absorbing liquor in slurry form. The solids in the absorbing liquor are mainly magnesium sulfite, magnesium sulfate, and magnesium oxide (insoluble). MgO slurry is added as makeup to the venturi absorber to maintain the proper reagent concentration. The clean flue gas leaves the absorber system through a separator and is mixed with hot flue gas from the dryer and discharged to the atmosphere through a stack.

A small stream of liquor from the absorption system is pumped to a centrifuge in which $MgSO_3$ and $MgSO_4$ crystals and unreacted MgO are separated from the mother liquor which is returned to the absorption system. It is desired to produce a cake with minimum moisture content. The centrifuged wet cake is dried in a direct-fired drier to remove the moisture. Anhydrous dry crystals from the drier are then calcined (decomposed by heat) in a direct-fired calciner to regenerate magnesium oxide and sulfur dioxide. The regenerated MgO is returned to the MgO makeup system, to which makeup water is added to produce makeup slurry for the absorption system. Additional MgO is also added to the makeup tank to compensate for process losses.

The reactions (Ref. 14) taking place in the major components of the Chemico process are summarized as follows:

CHEMICO MgO SCRUBBER SYSTEM - OIL FIRED BOILERS

(Ref. Chemical Construction Corporation)

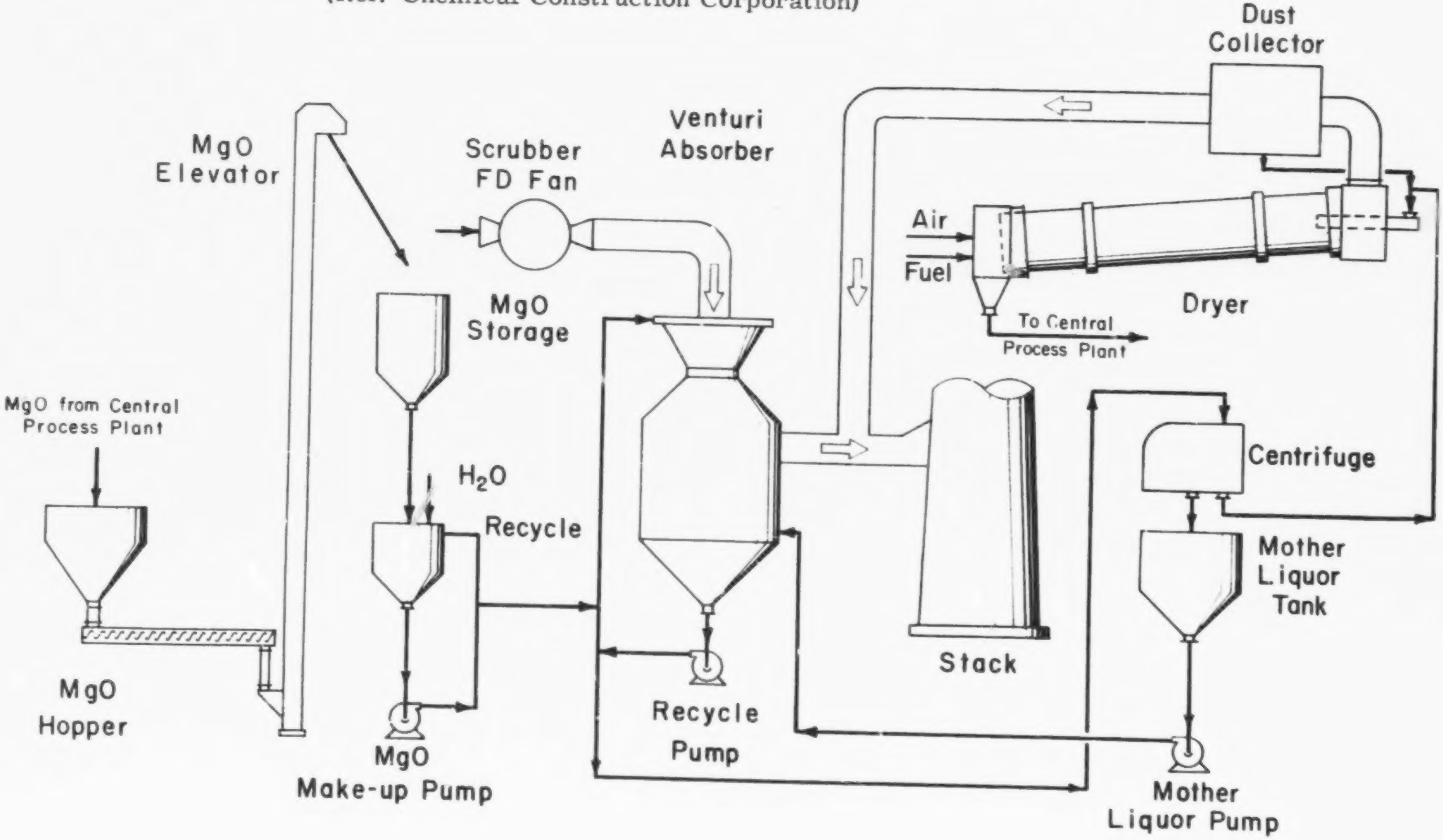

Dust Collector
MgO Elevator
Scrubber FD Fan
Venturi Absorber
Air
Fuel
Dryer
To Central Process Plant
MgO Storage
MgO from Central Process Plant
H$_2$O
Recycle
Centrifuge
Mother Liquor Tank
MgO Hopper
Stack
Recycle Pump
MgO Make-up Pump
Mother Liquor Pump

Figure IV-3

Main Reaction:

$$MgO + SO_2 + 6H_2O \longrightarrow MgSO_3 \cdot 6H_2O$$

Side Reactions:

$$MgSO_3 + SO_2 + H_2O \longrightarrow Mg(HSO_3)_2$$

$$Mg(HSO_3)_2 + MgO \longrightarrow 2MgSO_3 + H_2O$$

$$MgO + SO_3 + 7H_2O \longrightarrow MgSO_4 \cdot 7H_2O$$

$$2MgSO_3 + O_2 + 14\ H_2O \longrightarrow 2(MgSO_4 \cdot 7H_2O)$$

Drier:

$$MgSO_3 \cdot 6H_2O \xrightarrow{\text{Heat}} MgSO_3 + 6H_2O \uparrow$$

$$MgSO_4 \cdot 7H_2O \xrightarrow{\text{Heat}} MgSO_4 + 7H_2O \uparrow$$

$$H_2O \xrightarrow{\text{Heat}} H_2O \uparrow$$

Calciner:

$$MgSO_3 \xrightarrow{\text{Heat}} MgO + SO_2 \uparrow$$

$$2MgSO_4 + C \xrightarrow{\text{Heat}} 2MgO + 2SO_2 \uparrow + CO_2 \uparrow$$

A schematic of the magnesia process proposed by Babcock and Wilcox is shown in Figure IV-4 (Ref. 15). This configuration is assumed to be integrated into an 800 Mw boiler system.

The system starts at the plenum which collects the flue gas leaving the primary and secondary air heaters. The flue gas passes through the following equipment which is arranged in parallel: four venturi particulate scrubbers, four sulfur dioxide absorbers, four induced draft fans, and two steam coil gas reheaters. The gas is then discharged through a stack. The entrances to the stack are considered terminal points.

The gas entering the venturis is mixed with water which contacts and removes the entrained fly ash. The resulting fly ash slurry is transferred to a thickener for separation. The clarified overflow is returned to the venturis for reuse while the underflow is pumped to filters. Here, the fly ash is washed with fresh water to minimize the amount of soluble sulfur and magnesium compounds that leave the system with the fly ash. Pumps

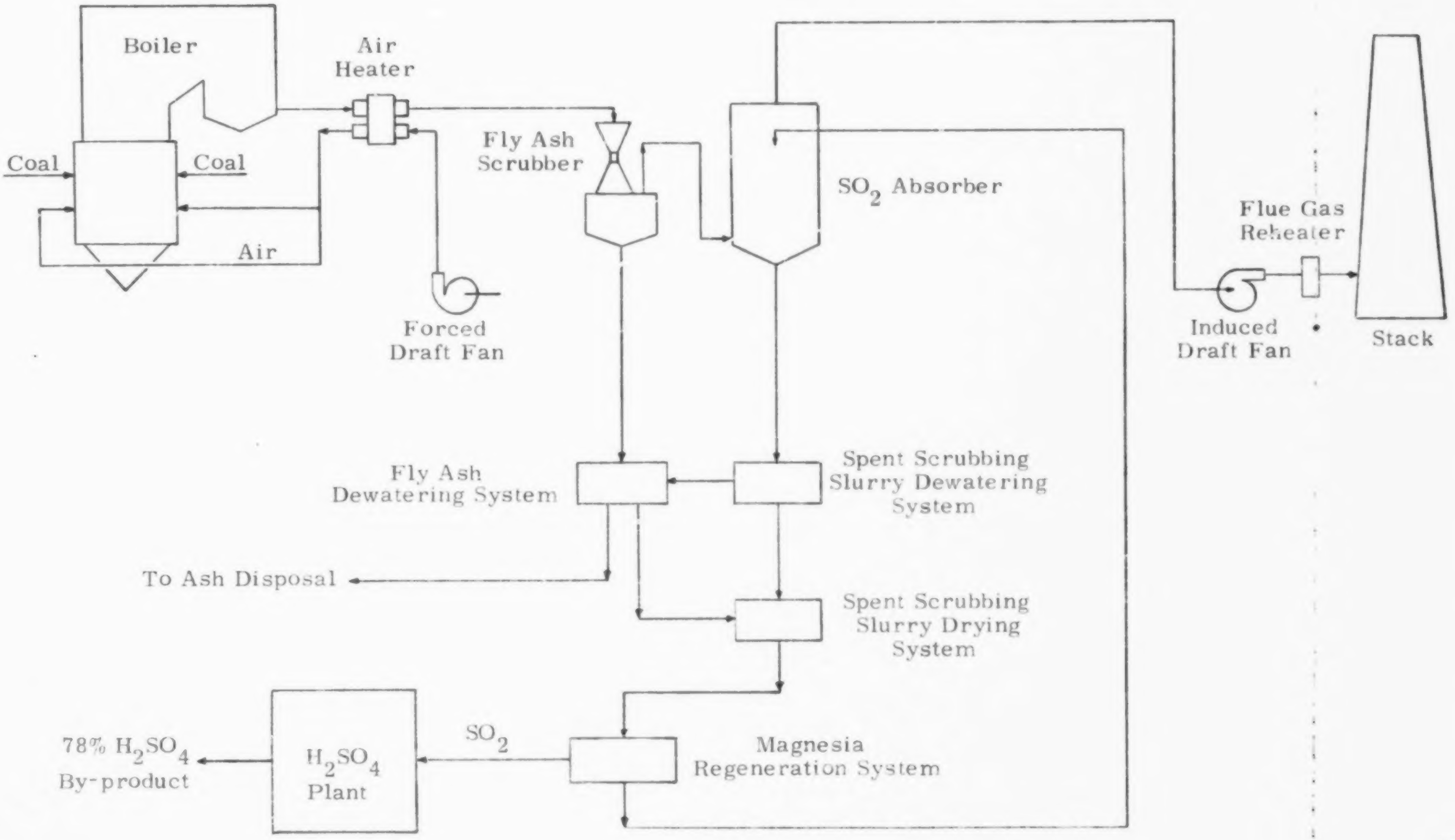

Boiler
Air Heater
Coal
Coal
Air
Forced Draft Fan
Fly Ash Scrubber
SO_2 Absorber
Flue Gas Reheater
Induced Draft Fan
Stack
Fly Ash Dewatering System
Spent Scrubbing Slurry Dewatering System
To Ash Disposal
Spent Scrubbing Slurry Drying System
78% H_2SO_4 By-product
H_2SO_4 Plant
SO_2
Magnesia Regeneration System

Figure IV-4. Babcock & Wilcox MgO Process

have been included to sluice the fly ash to a pond. The discharge of these pumps is the terminal point of the ash system.

The particulate-free flue gas passes to the absorbers where sulfur dioxide is removed by reacting with magnesium sulfite. The resulting magnesium-sulfur compounds are dewatered by thickening and centrifuging. The liquor which is separated from the slurry in the thickener and centrifuges is returned to the absorbers. Magnesium oxide, which results from the regeneration of the magnesium-sulfur compounds, is added to the absorbers to replace the magnesium which reacts with sulfur dioxide and is removed to the thickener as a sulfide. The solid materials leaving the centrifuge are dried and then decomposed to yield magnesium oxide and a gas stream containing sulfur dioxide. The product gas leaving the reactor is cooled, filtered, and leaves the system. It is then converted into sulfuric acid at an acid plant.

In the pilot plant studies performed by Babcock and Wilcox (Ref. 16) the efficiencies of the venturi-type and tray-type absorbers were measured. These absorber types are employed in the Chemico system and the Babcock and Wilcox system, respectively. The SO_2 removal efficiency for the venturi absorber is presented in Figure IV-5 as a function primarily of venturi pressure drop. It was reported that, except for pressure drop, the SO_2 removal performance of the venturi absorber showed a minimal dependence upon the parameters examined. The other variables considered were liquid-to-gas ratio, MgO/SO_2 stoichiometric ratio, pH of the scrubbing slurry, the presence of fly ash, the presence of moderate concentrations of $MgSO_4$, slaking or conversion of MgO to $Mg(OH)_2$ before injection into the venturi, and residence time of the liquid. The correlation between efficiency and pressure drop in the figure was judged as being based on insufficient data since there were only three data points at elevated pressure drop.

The SO_2 removal efficiency for the tray-type absorber or the floating bed absorber is shown in Figure IV-6. The SO_2 absorption data correlated well in terms of two operating parameters, L/G and spray slurry composition. The slurry composition,which is dependent upon the ratio of MgO/SO_2 and solubility of $MgSO_3$, can be expressed indirectly in terms of the bisulfite concentration, $Mg(HSO_3)_2$ or slurry pH. Thus, the figure illustrates the absorption efficiency as a function of L/G and bisulfite concentration.

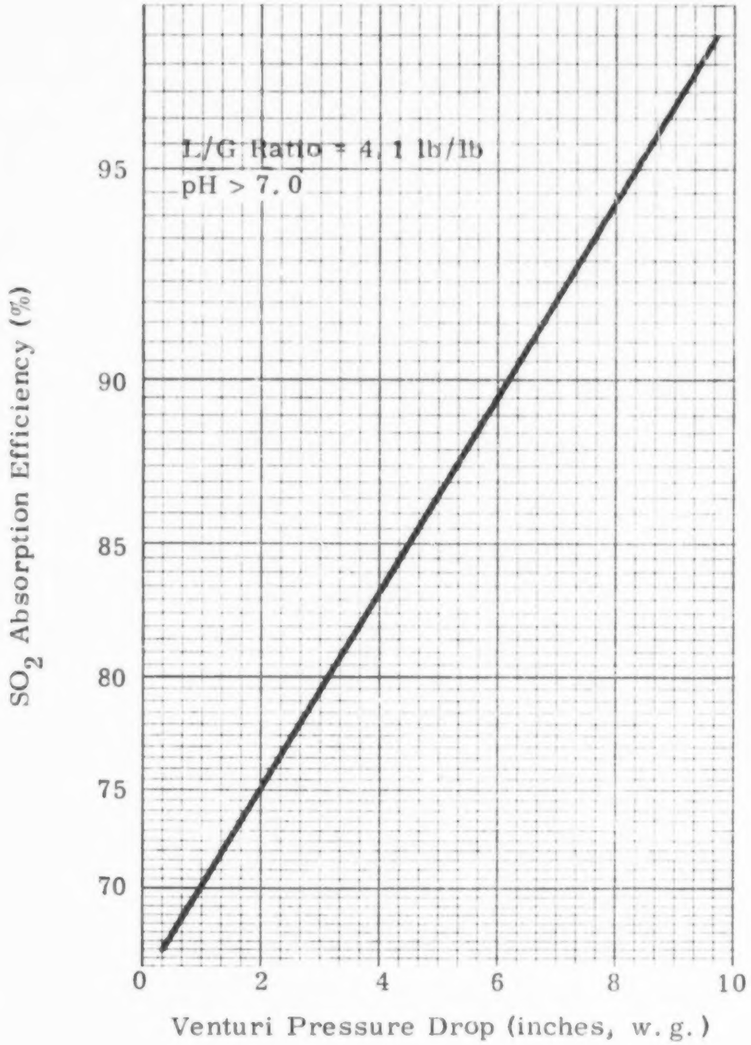


Figure IV-5. SO$_2$ Removal Efficiency for Venturi-Type Absorber

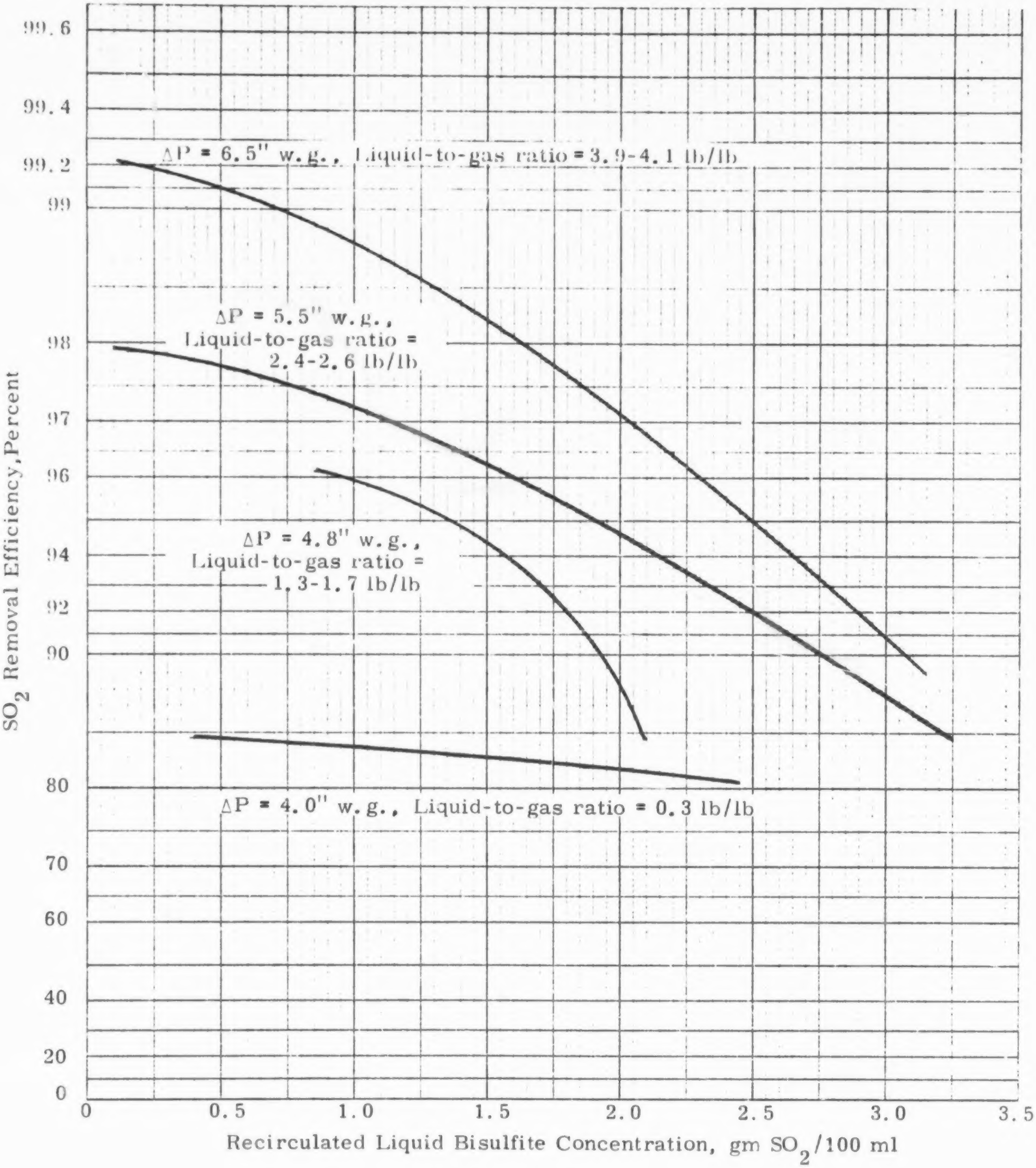


Figure IV-6. SO_2 Removal Efficiency for Floating Bed Absorber

c. Process Economics. The variations in capital and operating costs for the magnesium oxide are illustrated on nomograph pages 14-16. Each of the figures included in the nomograph is discussed in the succeeding paragraphs.

The five fixed capital cost estimates performed by the Kellogg Company apply to a Babcock and Wilcox system. Major equipment costs for this system are contributed by the venturi particulate scrubbers, tray-type absorbers, centrifuges, driers, gas reheaters, calciners, and the H_2SO_4 plant. An estimate was also performed by Hittman Associates at a plant capacity of 2000 Mwe and a 3.5 weight-percent sulfur coal. The resulting fixed capital cost of $42.8 million was an extension of the base case computed by Kellogg. One additional cost coordinate occurs for the trivial case at zero fixed capital cost and flow rate. Additional constant sulfur curves were obtained by interpolation and extrapolation of the data so that the range of sulfur content in fuels from 0.5 through 5.0 $lb/10^6$ Btu was covered. The final curves which were plotted on log-log graph paper allows one to obtain a fixed capital cost at a given flow rate and fuel sulfur content. The fixed cost in millions of dollars per year is then obtained for the fixed capital cost and a predetermined fixed charge rate.

Raw material costs are due to the annual requirement for makeup magnesium oxide. Makeup MgO quantity depends upon losses taking place in the drier and calciner or regeneration operations of the system. Magnesia requirement also varies with plant capacity and energy utilization. Magnesium oxide quantity has been assumed to vary proportionally with energy utilization and is based on the annual requirements presented in the base case.

The by-product credit for 98.5 percent concentrated H_2SO_4 is shown to be a linear function of sulfur removed. The sulfur removed by the process is the product of three quantities: SO_2 removal efficiency, energy utilization, and sulfur content. Appendix Table B-9 indicates that the assumption throughout of a constant SO_2 removal efficiency of 95 percent, exhibited by the three 3.5 percent sulfur cases, will introduce little error in the sulfur removed quantity. The quantity of 98.5 percent H_2SO_4 is then a constant multiple of the sulfur removed quantity.

Other operating costs include utilities, direct labor, supervision, maintenance, and overheads. These are presented as a function of energy utilization and the sulfur content of the fuel ranging from 0.5 through 5 $lb/10^6$ Btu.

4. Catalytic Oxidation Process

a. Process History. The catalytic oxidation process ("Cat-Ox") developed by Monsanto for the removal of sulfur oxides from stack gases is essentially an adaptation of the well-known contact process for the manufacture of sulfuric acid. The gas containing both sulfur dioxide and oxygen is passed through a fixed bed of catalyst at an appropriate temperature and most of the SO_2 is oxidized to SO_3. The gas is then passed through an absorption tower where the SO_3 is absorbed in recirculated sulfuric acid. The by-product for the Cat-Ox process is 78 percent concentrated sulfuric acid.

For power plant applications, the Cat-Ox process is offered in two configurations, one for use in new plants and one for previously existing plants. Several years of operating experience were gained on a now-dismantled pilot plant at Metropolitan Edison's Portland, Pennsylvania, station. A retrofitted system is planned for Illinois Power's 100 Mw Wood River 4 unit. Initial operation is expected in mid-1972.

b. Process Description. For the configuration that is integral with a new plant (Figure IV-7), components that are required in common for both the power plant and the Cat-Ox process are the electrostatic precipitator, fly ash handling system, economizer, air heater, steam/air heater, induced draft fan, and the forced draft fan. The electrostatic precipitator operates at a high temperature and is required to be highly efficient. In the converter, a single pass through the vanadium pentoxide catalyst bed results in 90 per cent conversion of SO_2 to SO_3. The air heater recovers less of the heat from the flue gas temperature than is done in a conventional power plant. The outlet flue gas temperature at the air heater is determined by the necessity for remaining above the acid dew point to avoid corrosion. Combustion air is preheated first in the fluid/air heater and second in the steam/air heater. The fluid/air heater is employed to recover additional heat from the recirculated acid which is cooled after each pass through the absorbing tower. The absorption of the sulfur trioxide in the cooled, recirculated acid is accomplished in a packed tower. The sulfuric acid mist formed during the cooling of the gas, plus any entrained droplets of the circulating acid carried out of the tower, are recovered in a fiber-bed mist eliminator. The minimum outlet gas temperature at the absorbing tower is set by the concentration of the sulfuric acid produced (78 percent).

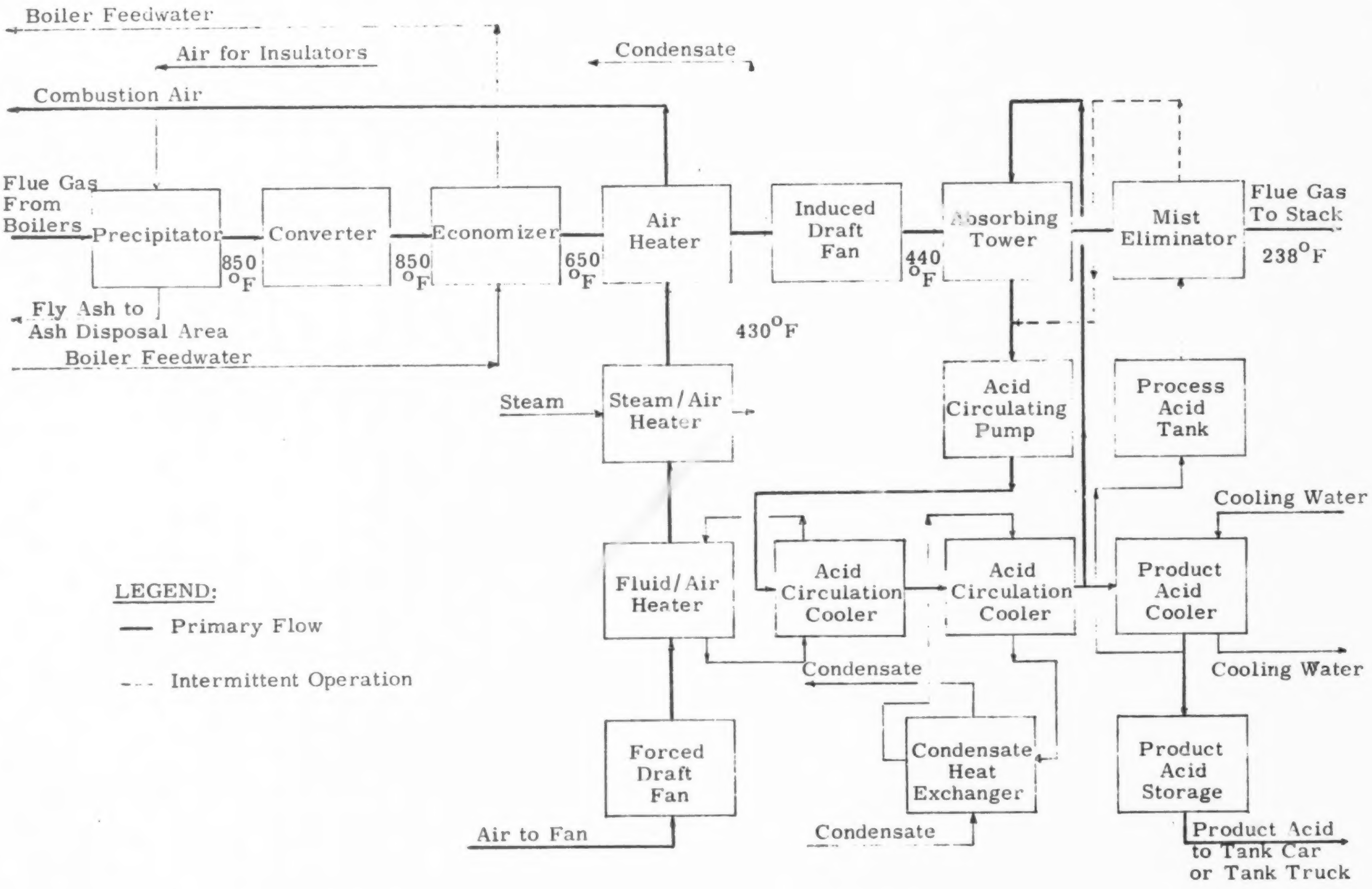

Boiler Feedwater
Air for Insulators
Combustion Air
Condensate
Flue Gas From Boilers
Precipitator
850 °F
Converter
850 °F
Economizer
650 °F
Air Heater
Induced Draft Fan
440 °F
Absorbing Tower
Mist Eliminator
Flue Gas To Stack
238°F
Fly Ash to Ash Disposal Area
Boiler Feedwater
430°F
Steam
Steam/Air Heater
Acid Circulating Pump
Process Acid Tank
Cooling Water
Fluid/Air Heater
Acid Circulation Cooler
Acid Circulation Cooler
Product Acid Cooler
LEGEND:
Primary Flow
Intermittent Operation
Condensate
Cooling Water
Forced Draft Fan
Condensate Heat Exchanger
Product Acid Storage
Air to Fan
Condensate
Product Acid to Tank Car or Tank Truck

Figure IV-7. Integrated Catalytic Oxidation System For New Power Plant

The Cat-Ox reheat system for an existing power plant is shown schematically in Figure IV-8. It is additive instead of integral. The reheat system could be bypassed entirely without affecting the operation of the power plant except during a relatively short period in which the final tie-ins are made. This period could coincide with the annual shutdown of the power plant for overhaul.

The reheat system takes the gas from the discharge of the existing induced draft fan. A new precipitator is added to reduce the quantity of dust to a tolerable level, which is the same as in the integrated Cat-Ox system. A new induced draft fan is added to move the gas through the Cat-Ox system. The relatively cool (325°F) flue gas must be heated to a temperature high enough to permit catalytic oxidation of the sulfur dioxide. Part of this preheating is accomplished by exchange of heat from the converter exit gas in the gas heat exchanger. The rest of the preheating is accomplished by direct injection of hot combustion gases from the reheat furnace, which can be fired with No. 2 fuel oil of low ash content. The hot gas leaving the converter passes through the gas heat exchanger where it heats the incoming cool feed gas and is itself cooled to 450°F before it enters the absorbing tower. From this point on, the arrangement and operation are essentially the same as in the integrated Cat-Ox system. The temperature of the gas leaving the heat exchanger, 450°F, is set as in the integrated system by the necessity of staying above the sulfuric acid dew point. In the gas heat exchanger, as in the air heater of the integrated system, it is necessary to avoid cold-end corrosion. The temperature of the incoming cool flue gas must therefore be raised somewhat and this is accomplished by the recycle of a portion of the hot exit gas to the inlet of the exchanger. The sizes of the new induced draft fan and the gas heat exchanger must be increased accordingly to handle the greater total flow of gas. The acid circulated to the absorbing tower is cooled directly with cooling water and there is no recovery of the low-level heat. The product acid is of 78 percent concentration, as in the integrated system.

For a new or integrated system, SO_2 removal efficiency depends upon the converter temperature and the contact time in the catalyst bed. Reference 17 indicates that this efficiency is 90 percent at the nominal converter temperature of 850°F. For a retrofitted process or reheat system,

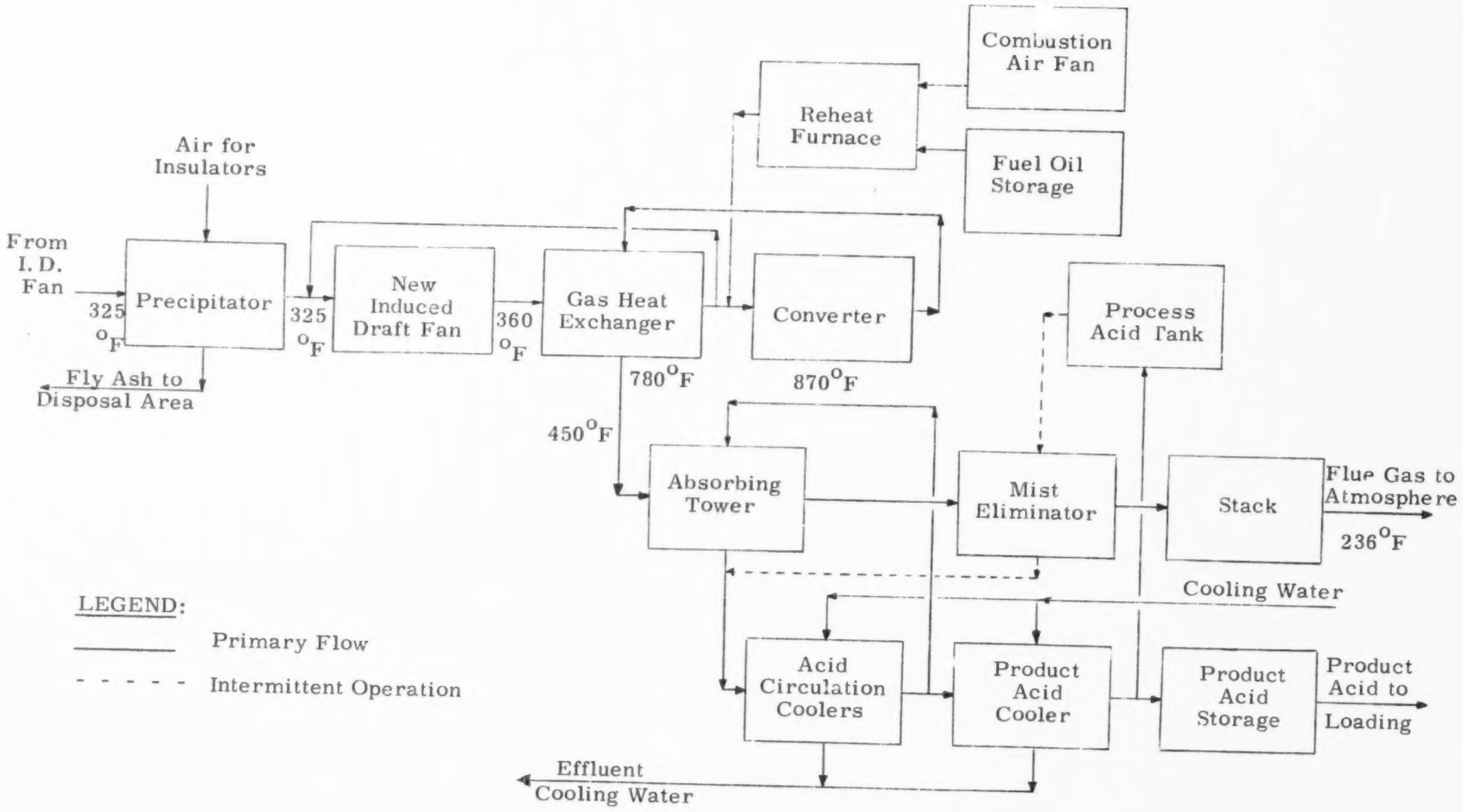

From I. D. Fan
325 °F
Air for Insulators
Precipitator
Fly Ash to Disposal Area
325 °F
New Induced Draft Fan
360 °F
Gas Heat Exchanger
780°F
450°F
Reheat Furnace
Combustion Air Fan
Fuel Oil Storage
Converter
870°F
Absorbing Tower
Mist Eliminator
Process Acid Tank
Stack
Flue Gas to Atmosphere
236°F
Cooling Water
Acid Circulation Coolers
Product Acid Cooler
Product Acid Storage
Product Acid to Loading
Effluent Cooling Water
LEGEND:
Primary Flow
Intermittent Operation

Figure IV-8. Catalytic Oxidation Reheat System for Existing Power Plant

an overall SO_2 removal efficiency of 90 percent is specified by Monsanto. Graphical presentations of the SO_2 removal efficiency are unavailable at this time for the Cat-Ox process.

c. Process Economics. The variations in capital and operating costs for the Cat-Ox process are illustrated by a series of figures on nomograph pages 17-19. The fixed capital cost estimates generated by the Kellogg Company apply to a new process or an integrated Cat-Ox system. Additional estimates were computed by Hittman Associates at plant capacities of 1500 and 2000 Mw and a 3.5 weight-percent sulfur coal. The resulting magnitudes were \$57.2 million and \$69.6 million, respectively. These were extensions of the base case. The family of fixed capital cost curves, spanning the range of sulfur content from 0.5 through 5.0 $lb/10^6$ Btu, is closer together than for the MgO process. The capital costs for the Cat-Ox process are thus less sensitive to variations in sulfur content found in fuels. A fixed cost in millions of dollars per year is obtained for a given fixed capital cost and a given fixed charge rate.

Raw material costs are due to the annual requirement for vanadium pentoxide (V_2O_5). V_2O_5 replacement cost is incurred by a loss of the catalyst. This is caused by the cleaning operations necessitated by the gradual buildup of fly ash in the catalyst bed. Even though an electrostatic precipitator with 99.6 percent efficiency is used, enough fly ash enters the bed to require four cleanings per year. A 2.5 percent loss of catalyst occurs during each cleaning. The replacement quantity is dependent upon the energy utilization and the fly ash content of the fuel. Since the exact functional dependency is unknown, the base case operating cost estimate was used to establish a family of 45^o lines for the annual volume of V_2O_5. The ash content of the coal on the as-received basis was 14.5 percent.

The by-product credit for 78 percent concentrated H_2SO_4 is shown as a function of sulfur removed. The sulfur removed by the Cat-Ox process is presented versus energy utilization and sulfur content after assuming an SO_2 removal efficiency of 90 percent. Table B-10 of Appendix B indicates that this assumption will introduce a maximum error of under 10 percent in the sulfur removed quantity. The quantity of 78 percent H_2SO_4 is then a constant times the sulfur removed quantity.

Other operating costs include utilities, direct labor, supervision, maintenance, and overheads. A credit for stack gas cooling has been accounted for in the process. Other operating cost exhibits negligible variation with changing sulfur content. Thus, a single line results after plotting other operating costs versus energy utilization.

5. Modified Chamber Process

a. Process History. The catalytic chamber process enables the removal of both the oxides of sulfur and nitrogen from power plant flue gas. The process is a modified version of an earlier chamber method to produce sulfuric acid. Commercial grades of sulfuric and nitric acids are by-products of the process. The development of the process was initiated in October 1967 and proceeded until October 1970 under two contracts for the Environmental Protection Agency. The effort by Tyco Laboratories, Inc., is continuing under a new contract to the Air Quality Office. To date, the experimental work has consisted of a laboratory scale investigation of the separate process stages and a process evaluation on a continuous basis in a 10 scfm pilot plant.

b. Process Description. A schematic diagram of the Tyco catalytic chamber process (Ref. 18) is shown in Figure IV-9. The flue gas, fed to the process, is mixed with recycled nitrogen dioxide which oxidizes the sulfur dioxide to sulfur trioxide in a reactor. Water vapor in the flue gas stream converts the SO_3 to sulfuric acid. The reacted gases pass through the scrubber which removes the newly formed acids and the oxides of nitrogen while venting the cleaned, wet stack gas to the atmosphere. Fly ash scrubbed from the flue gas is filtered from the acid stream. The nitrogen oxide-bearing sulfuric acid is then passed onto the catalytic stripper where it is contacted with air in the presence of activated carbon. Here the nitrosylsulfuric acid is catalytically oxidized to H_2SO_4 and NO_2. The NO_2-bearing gas stream is fed to a nitric acid production stage where the excess NO_2 is recovered. The H_2SO_4 leaving the stripper is recycled to the scrubber after the by-product acid is removed. This by-product acid is equivalent to the amount of SO_2 recovered from the flue gas in the scrubber after oxidation by NO_2.

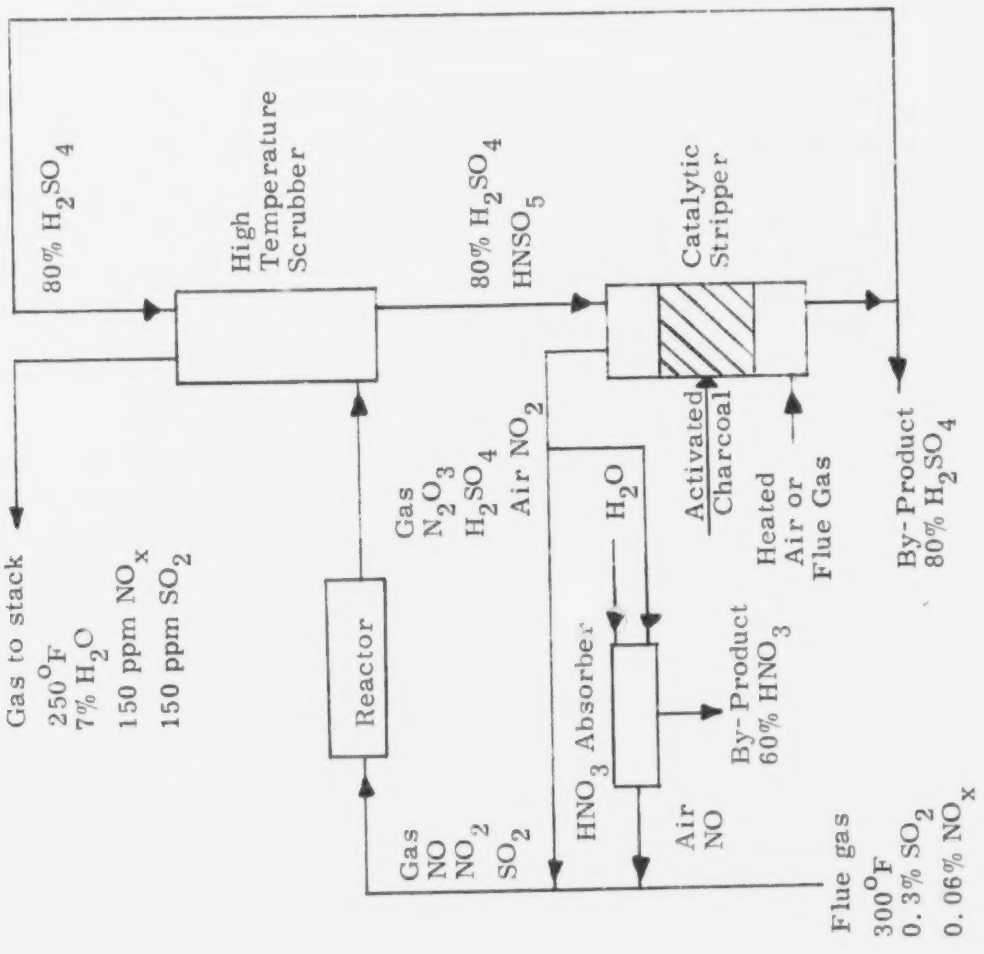

Gas to stack
250°F
7% H_2O
150 ppm NO_x
150 ppm SO_2
80% H_2SO_4
High Temperature Scrubber
Reactor
Gas
NO
NO_2
SO_2
Gas
N_2O_3
H_2SO_4
Air NO_2
80% H_2SO_4
$HNSO_5$
Catalytic Stripper
HNO_3 Absorber
H_2O
Air
NO
By-Product
60% HNO_3
Activated Charcoal
Heated Air or Flue Gas
By-Product
80% H_2SO_4
Flue gas
300°F
0.3% SO_2
0.06% NO_x

Figure IV-9. Tyco Catalytic Chamber Process Flow Diagram

The reactions taking place in the major components of the process are summarized as follows:

Reactor:

$$SO_2 + NO_2 \rightleftharpoons SO_3 + NO$$

$$SO_3 + H_2O \rightleftharpoons H_2SO_4$$

$$2NO + O_2 \rightleftharpoons 2NO_2$$

$$NO + NO_2 \rightleftharpoons N_2O_3$$

Scrubber:

$$N_2O_3 + 2H_2SO_4 \rightleftharpoons 2HNSO_5 + H_2O$$

Catalytic Stripper:

$$4HNSO_5 + O_2 + 2H_2O \rightleftharpoons 4NO_2 + 4H_2SO_4$$

Nitric Acid Production:

$$3NO_2 + H_2O \longrightarrow 2HNO_3 + NO$$

The SO_2 removal efficiency for this process is a function of reactor residence time, reactor temperature profile, and the NO_2:SO_2 ratio into the reactor. From Reference 18, following experimentation with the SO_2 reactor, a residence time of approximately 15 seconds was required to yield an SO_2 removal efficiency of 95 percent from flue gas containing 0.3 percent SO_2 by volume. The NO_2:SO_2 ratio into the reactor is nominally 2.0, but it can range from less than 2.0 up to 2.2. The reactor temperature profile used in the pilot plant was relatively flat since the temperature difference between inlet and exit was small, 335°F and 372°F, respectively (Ref. 19). Thus, the reactor operating temperature used in obtaining the data was approximately constant. Following further operation of the 10 scfm miniplant, a better definition should result for the parametric variation of SO_2 removal efficiency.

c. Process Economics. The variations in capital and operating costs for the modified chamber process are illustrated by a series of figures on nomograph pages 20 22. The fixed capital cost estimates generated by the Kellogg Company apply to a new Tyco process employing a packed scrubber as opposed to a plate tower scrubber. Major equipment costs for this system are contributed by the flue gas blowers, the reactors, the packed scrubbers, and the catalytic strippers. An additional capital cost estimate was computed by Hittman Associates as being \$48.3 million at a plant capacity of 2000 Mw and a 3.5 weight-percent sulfur coal.

Fixed capital cost versus flue gas flow rate is represented by a single line. The costs exhibited negligible variations resulting from perturbations in sulfur content. The annual fixed cost in millions of dollars per year depends upon the capital cost from the first portion of the graph and the fixed charge rate.

Raw material costs are due to the replacement costs for the activated charcoal bed in the catalytic stripper and the limestone required to neutralize the filter cake. The parametric study of Reference 19 showed that the annual quantity of granular charcoal catalyst varied in direct proportion to energy utilization. It was not indicated as being a function of the fuel sulfur content. Therefore, the annual cost to replace the charcoal beds is shown parametrically to account for variations in unit cost alone. The annual cost of limestone is small (approximately one percent of direct costs) and consequently is lumped into "other operating costs."

Credits are obtained for the by-product acids, 80 percent concentrated H_2SO_4 and 60 percent concentrated HNO_3. It was shown in Reference 19 that the annual quantity of sulfuric acid was a function of energy utilization and sulfur content of the fuel. The quantity is shown as varying proportionately to increases in energy utilization and variations in sulfur content. Therefore, the annual revenue for by-product H_2SO_4 is obtained for a given energy utilization, sulfur content, and net sales revenue. The sulfur removed at an SO_2 removal efficiency of 87.5 percent can also be obtained from the graph.

The annual quantity of by-product nitric acid is a function of energy utilization and the NO_x content of the flue gas. The HNO_3 quantity is proportional to energy utilization. This quantity was shown (Ref. 19) not to be directly proportional to the NO_x content of the flue gas. The concentration of nitrogen oxides in the flue gas is a function of the boiler design and operating temperature, excess air, fuel type, and firing methods. For coal-fired units, a representative range of NO_x content in the flue gas by volume would be 400 to 1200 ppm. The Kellogg base case assumed a value of 500 ppm. This appears to be reasonable when compared to the magnitudes presented in Reference 20.

Other operating costs include utilities, direct labor, supervision, maintenance, overheads, and limestone material. A credit for boiler feedwater heating has also been accounted for. Other operating cost exhibits insignificant variation with changing sulfur content. Thus, a single line results after plotting it versus energy utilization.

6. Molten Carbonate Process

a. Process History. The molten carbonate process to remove sulfur oxides from power plant flue gases is being developed by the Atomics International (AI) Division of North American Rockwell. The effort was initiated on June 1, 1967, under contract with the National Air Pollution Control Administration of the Department of Health, Education, and Welfare. Bench studies and conceptual testing for the molten carbonate process have been performed. However, construction of a pilot plant for the process is only in the planning stages. During March 1969, an advisory committee of the National Academy of Engineering recommended that the pilot plant be built to treat a side stream of gases from an existing coal-burning boiler. It is anticipated that a site may soon be selected.

b. Process Description. In the process shown in Figure IV-10, flue gas containing sulfur oxides is scrubbed in absorbers with a molten mixture of lithium, potassium, and sodium carbonates (M_2CO_3). The process temperature is nominally 850°F. The melt, containing sulfite salts, is regenerated by heating to 1500°F where the sulfates, formed by disproportionation of the sulfites, can be reduced to sulfides with coke. The sulfides in the melt are converted back to carbonates, with the formation of hydrogen sulfide, by reaction at 850-900°F with carbon dioxide and steam. The H_2S is sent to a Claus Plant for recovery of by-product sulfur. A description follows of the reactions taking place in the major components of the process.

Flue gases tapped before the economizer section of the boiler are passed through a bank of high temperature electrostatic precipitator sections operating in parallel. Precipitator efficiency is 99.5 percent. The gases are sent from each precipitator to a corresponding absorber for removal of the sulfur oxides by the molten carbonate stream. The return flue gas to the economizer section of the boiler has had greater than 90 percent of the original SO_x removed.

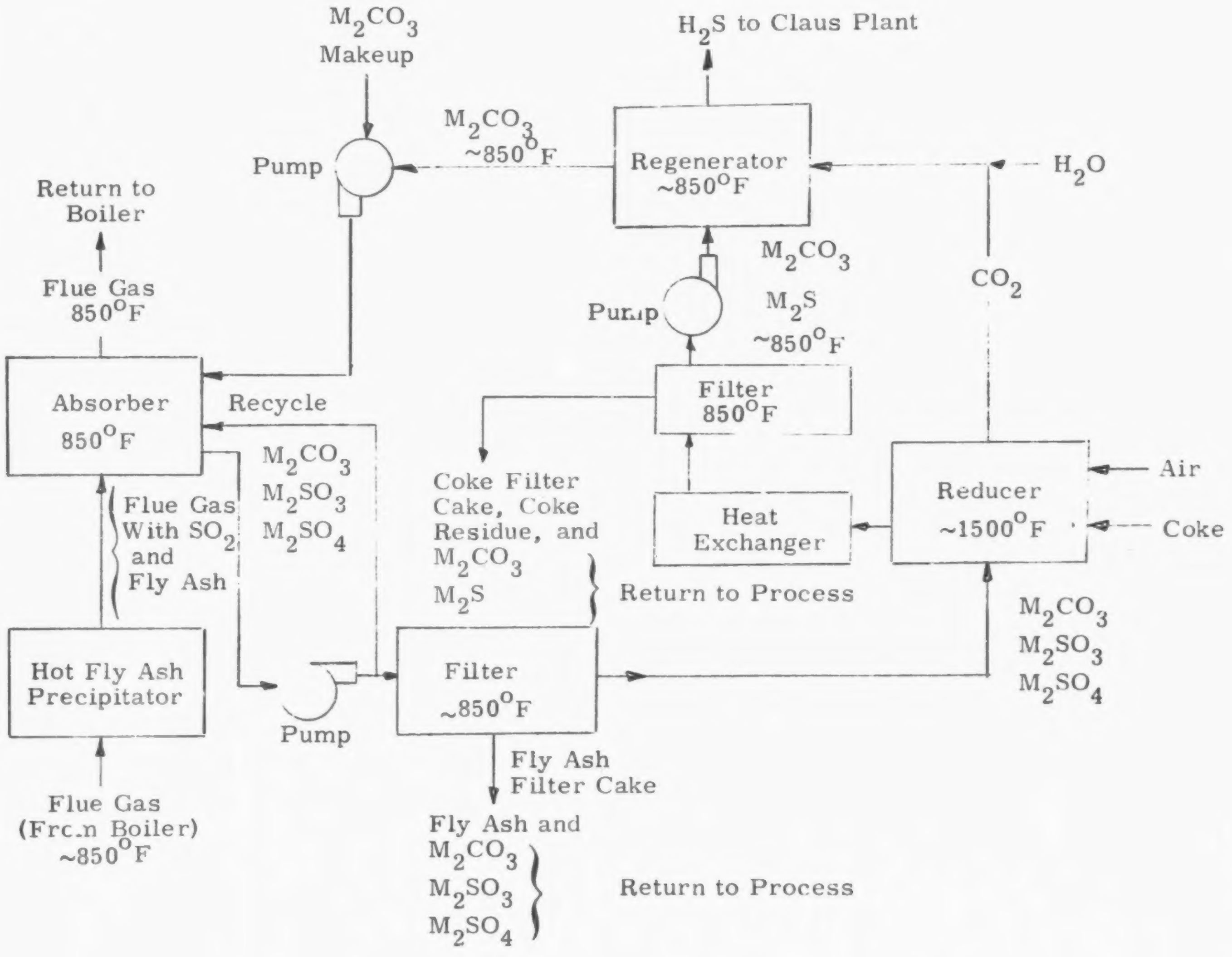

M_2CO_3 Makeup
H_2S to Claus Plant
Pump
M_2CO_3 ~850°F
Regenerator ~850°F
H_2O
Return to Boiler
M_2CO_3
Flue Gas 850°F
CO_2
Pump
M_2S
~850°F
Absorber 850°F
Recycle
Filter 850°F
M_2CO_3
M_2SO_3
M_2SO_4
Reducer ~1500°F
Air
Coke
Flue Gas With SO_2 and Fly Ash
Coke Filter Cake, Coke Residue, and
M_2CO_3
M_2S
Heat Exchanger
Return to Process
M_2CO_3
M_2SO_3
M_2SO_4
Hot Fly Ash Precipitator
Filter ~850°F
Pump
Fly Ash Filter Cake
Flue Gas (From Boiler) ~850°F
Fly Ash and
M_2CO_3
M_2SO_3
M_2SO_4
Return to Process

Figure IV-10. Molten Carbonate Process Flow Diagram

Absorber reactions are:

$$SO_2 + M_2CO_3 \longrightarrow M_2SO_3 + CO_2$$

$$SO_3 + M_2CO_3 \longrightarrow M_2SO_4 + CO_2$$

where M represents the cutectic mixture of sodium, potassium, and lithium cations. Some of the M_2SO_3 may be oxidized by oxygen in the flue gas as follows:

$$2M_2SO_3 + O_2 \longrightarrow 2M_2SO_4$$

All sulfides are also oxidized as follows:

$$M_2S + 2O_2 \longrightarrow M_2SO_4$$

Fly ash is removed from the process by two filter units in parallel. One unit is in operation while the other unit is being drained and a "dry" ash cake is discharged. The fly ash filter cake is composed of a mixture of molten salts and solids consisting of KCl, fly ash, and a trace of coke. The fly ash filter cake is sent to the Li_2CO_3 recovery process which recovers over 80 percent of the Li_2CO_3. At every discharge, a small amount of reactant is lost requiring a carbonate make-up.

Each reducer consists of oxidation and reduction zones with liquid passages between the two compartments for internal melt recirculation. The endothermic heat for reduction and the sensible heat required to raise the melt, coke, and air to the reduction temperature of 1500°F are provided by the oxidation of M_2S in the oxidation zone of the reducers. At 1500°F, the M_2SO_3 in the melt disproportionates into M_2SO_4 and M_2S. The M_2SO_4 is reduced in the reduction zone to M_2S by the carbon contained in the coke. Part of the reduction zone melt is recycled to the oxidation zone to oxidize the M_2S and provide the necessary heat. The oxidation zone melt also recycles back to the reduction zone to convert the M_2SO_4 into M_2S. Reducer reactions are as follows:

Oxidation:

$$M_2S + 2O_2 \longrightarrow M_2SO_4$$

$$C + O_2 \longrightarrow CO_2$$

Reduction:

$$M_2SO_4 + 2C \longrightarrow M_2S + 2CO_2$$

Disproportionation:

$$4M_2SO_3 \longrightarrow 3M_2SO_4 + M_2S$$

Coke Sulfur Recovery:

$$2S + C + 2M_2CO_3 \longrightarrow 2M_2S + 3CO_2$$

The melt from the reducer at 1500°F is cooled.

Melt from the reduction step flows to one of two parallel coke filters where the unreacted coke is removed from the process melt. The filtered melt is sent to the regenerator. The coke filter cake is composed of coke plus traces of KCl and fly ash. The coke filter cake is sent to the Li_2CO_3 recovery process. Additional carbonate make-up is needed to replace the nonrecovered carbonates in the coke filter cake.

The regenerator is a tower consisting of bubble cap trays. The melt is added to the top of the column and is contacted with the CO_2 and H_2O rising up the tower. The regeneration is exothermic and coolers are provided to remove this heat. The regeneration reaction is:

$$M_2S + CO_2 + H_2O \longrightarrow M_2CO_3 + H_2S$$

The gases from the regenerator containing H_2O, CO_2, and H_2S are sent to a Claus Plant for recovery of by-product sulfur

The absorber design criteria proposed by AI is incorporated into Figure IV-11 as presented by Singmaster and Breyer (Ref. 21). In addition, a maximum gas velocity of 25 fps is proposed. In the figure, the cross-hatched area represents the limiting region for operation. On the abscissa is the inlet molar ratio defined to be the total mols of molten carbonate per hour to the absorber divided by the mols of SO_x per hour entering in the flue gas. On the ordinate is the molten carbonate concentration at the absorber outlet. Recycle is defined on the figure. The desired target operating condition has the coordinates corresponding to a recycle of 1:1 and 68 percent carbonate melt concentration at the absorber outlet.

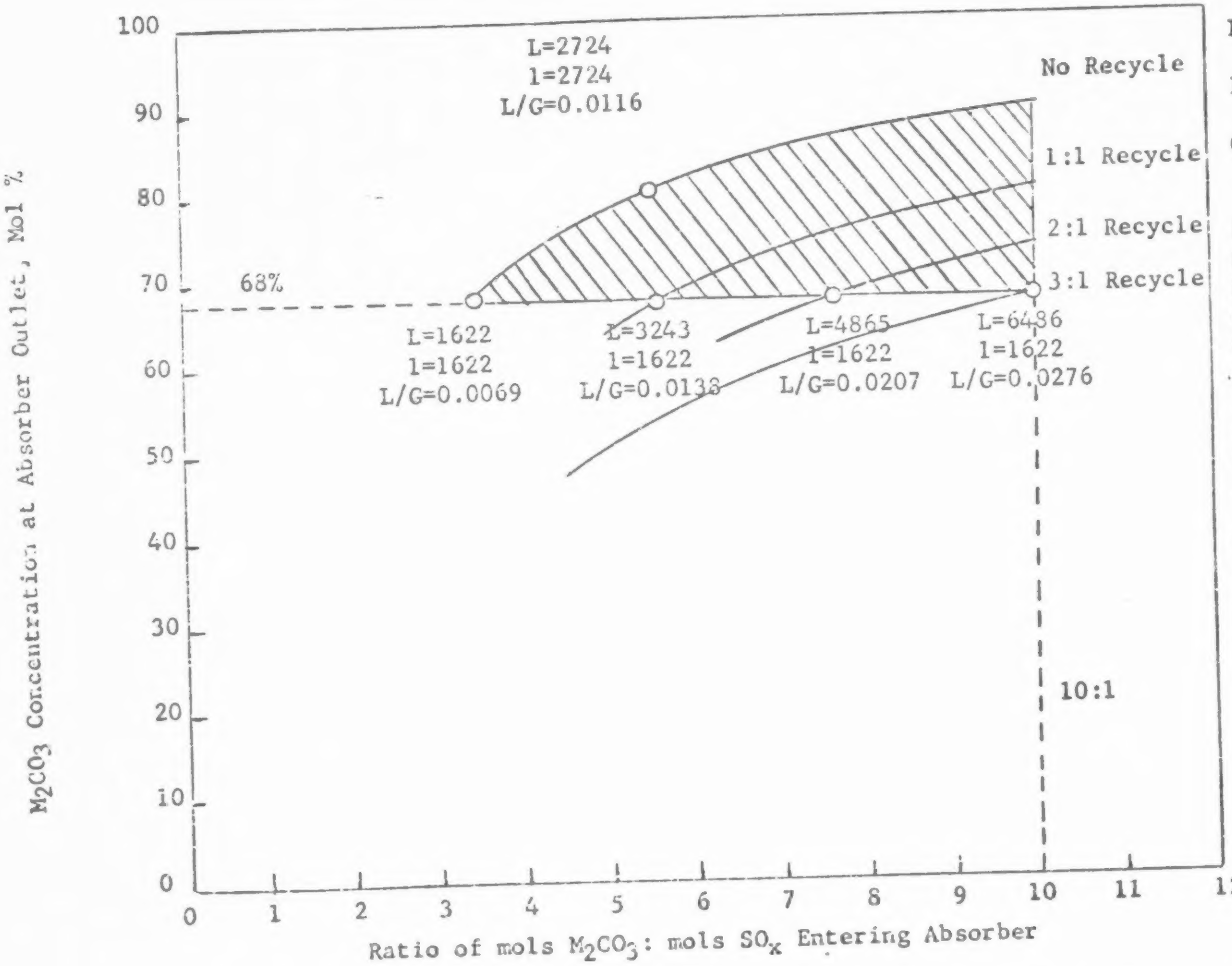


Figure IV-11. Absorber Operating Limits (Based on A-I Design Parameters)(21)

Design range is shown cross hatched.

L is total mols of melt to absorber, per hour.
1 is total mols of melt to reducer, per hour.
G is total mols of gas to absorber, per hour.

Recycle is the number of mols of melt which is returned to the absorber divided by the number of mols of melt advancing to the reducer.

Flow rates pertain to the case: 800 Mw, 3% S in coal.

Dissolved KCl not included in melt compositions.

The bounds for the inlet molar ratio are approximately 3.5:1 and 10:1 from the figure. In Reference 22, the SO_2 removal efficiency was presented versus the inverse of the inlet molar ratio. It is shown here in Figure IV-12. These data resulted from experiments employing an absorption column apparatus wherein operation occurred at low and high (fewer cases) gas velocities The efficiencies corresponding to these bounds were approximately 93 percent and 98 percent, respectively. Therefore, the SO_2 removal efficiency is very high and is essentially invariant over the range of operating conditions for the process absorber.

c. Process Economics. The variation in capital and operating costs for the molten carbonate process are illustrated by the series of figures on nomograph pages 23-26. The fixed capital cost estimates generated by the Kellogg Company applied to a new process. Additional estimates were computed by Hittman Associates at 1500 and 2000 Mw. Major equipment costs for this system are contributed by the high efficiency (η = 99.5 percent), high temperature electrostatic precipitator, the absorbers, the reducers, and the Claus sulfur plant. A family of curves was obtained for the range of sulfur content from 0.5 through 5.0 lb/10^6 Btu. A fixed cost is obtained for a given fixed capital cost and fixed charge rate.

The by-product credit for sulfur is determined after multiplying the unit cost of sulfur by the annual quantity of by-product sulfur. The by-product sulfur is known to be approximately equal to the quantity of sulfur removed from the flue gas. The quantity of sulfur removed was determined versus energy utilization and sulfur content at an SO_2 removal efficiency of 95 percent.

Raw material costs are due to coke and carbonate makeup requirements. The annual quantity of coke required as input to the reducer is directly proportional to the sulfur removed. The annual coke cost is read directly for a given sulfur removed and coke unit cost. If the unit cost of coke becomes excessive, alternate low cost raw materials may suffice as sources of the carbon required for reduction.

The annual replacement cost for makeup carbonate is determined from nomograph pages 25 and 26 by summing the following three contributions, respectively:

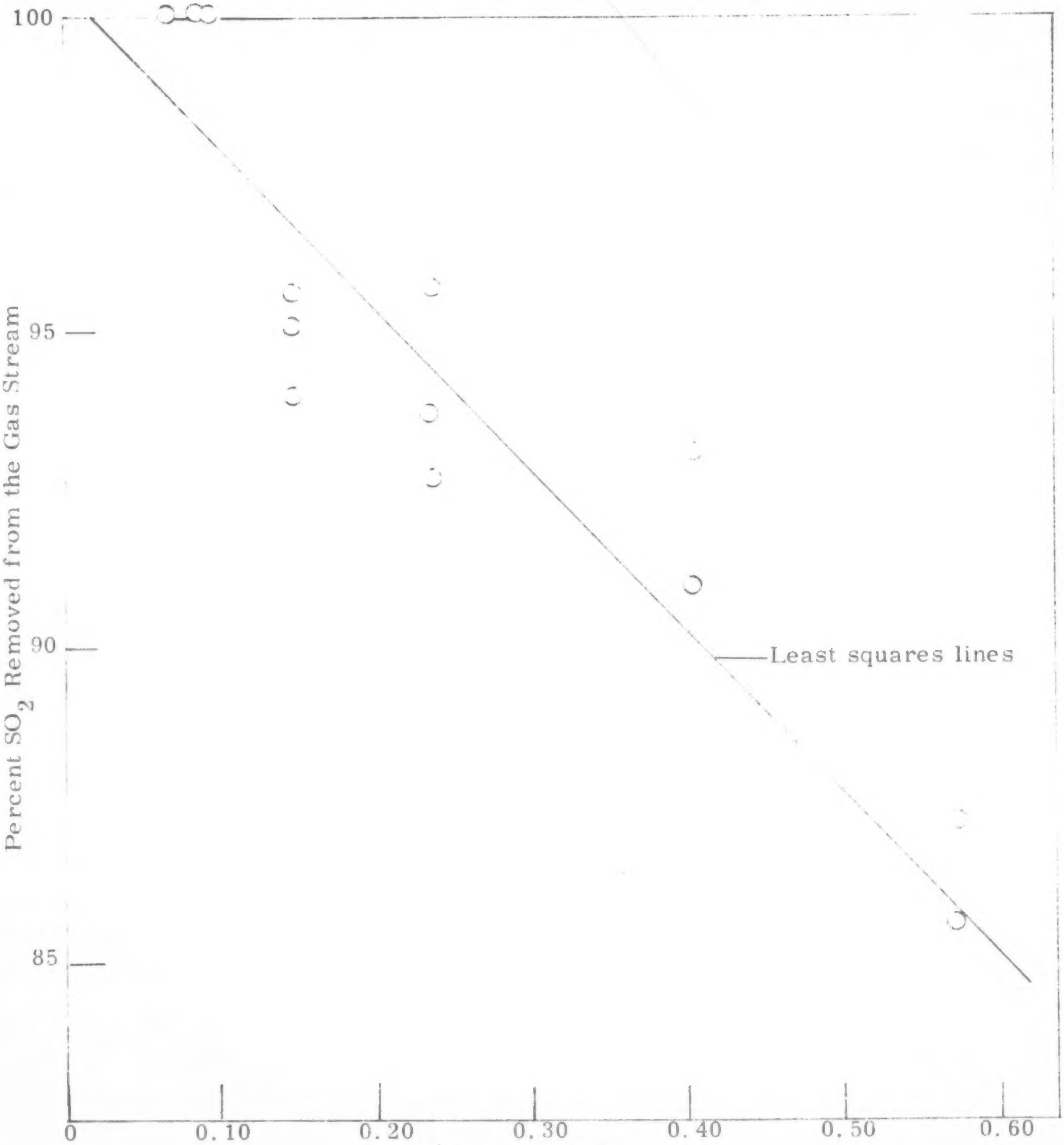


Figure IV-12. Effect of SO_2-M_2CO_3 Ratio Upon the Percent Removal of SO_2

(1) Makeup cost for M_2CO_3 loss due to fly ash filter cake

(2) Makeup cost for M_2CO_3 loss due to coke filter cake

(3) Makeup cost for M_2CO_3 loss due to KCl filter cake and KCl production

These costs depend upon energy utilization and sulfur removed (for the coke filter case case only) shown on the abscissa.

In Reference 21, Singmaster and Breyer developed the carbonate makeup costs shown in Table IV-3 for their base case, 800 Mw capacity. Thus, for a given energy utilization, it can be shown that the makeup cost for loss due to the fly ash filter cake is:

Cost contribution (1) = (Li_2CO_3 unit cost x lb of Li_2CO_3 makeup per lb of fly ash + K_2CO_3 unit cost x lb of K_2CO_3 makeup per lb of fly ash + Na_2CO_3 unit cost x lb of Na_2CO_3 makeup per lb of fly ash) x lb/hr of fly ash removed at fly ash filter x 8760 hr/yr x plant load factor

The Kellogg base case operating cost estimate of Table B-6 employed a fly ash removal rate corresponding to 14.5 percent ash content in the coal. The removal rate was increased and decreased proportionately in order to obtain makeup costs corresponding to the 5, 10, and 15 percent ash contents, respectively. Thus, makeup costs due to losses in the fly ash filter cake are presented as a function of energy utilization and the percent ash content in the coal.

Similarly, for a given energy utilization, the carbonate makeup cost for loss due to the coke filter cake is:

Cost contribution (2) = (Li_2CO_3 unit cost x lb of Li_2CO_3 makeup per lb of unreacted coke + K_2CO_3 unit cost x lb of K_2CO_3 makeup per lb of unreacted coke + Na_2CO_3 unit cost x lb of Na_2CO_3 makeup per lb of unreacted coke) x lb/hr of coke removed at the coke filter x 8760 hr/yr x plant load factor

TABLE IV-3. CARBONATE MAKEUP REQUIREMENTS ASSUMING 88 PERCENT LI_2CO_3 RECOVERY

(Ref. Singmaster and Breyer Base Case)

	Li_2CO_3	K_2CO_3	Na_2CO_3	TOTAL
Cost Per Pound	\$0.42	\$0.096	\$0.028	
Makeup Ratio Required for Non-Recovered Salt, #/# of				
Fly Ash and KCl Filter Cake	0.067	0.611	0.573	
Coke Filter Cake	0.043	0.368	0.361	
KCl Produced	-	0.926	-	
Annual Replacement for Loss From				
Fly Ash Filter Cake	93,000#	843,000#	791,000#	
KCl Filter Cake	167,000	1,510,000	1,417,000	
Coke Filter Cake	211,000	1,805,000	1,771,000	
KCl Production	-	2,288,000	-	
Total	471,000#	6,446,000#	3,979,000#	
Annual Cost for Makeup Due to Losses in				
Fly Ash Filter Cake	\$ 39,000	\$ 81,000	\$ 22,000	\$142,000
KCl Filter Cake	70,000	145,000	40,000	255,000
Coke Filter Cake	89,000	173,000	50,000	312,000
KCl Production	-	220,000	-	220,000
Total	\$198,000	\$619,000	\$112,000	\$929,000

Basis:

Filter Cake Constituent:
Fly ash-225 lb/hr
KCl -403 lb/hr
Removed at fly ash filter

Coke -800 lb/hr
Removed at coke filter

The unreacted coke or coke residue removal rate was determined parametrically in Reference 21 as being directly proportional to plant capacity and percent sulfur in the coal. The quantity in Table IV-3 is based on the assumption of complete oxidation of carbon in the coke to provide the reducer sensible heat and heat of reduction. A conversion of 95 percent of the M_2SO_4 to M_2S by carbon is also assumed. Therefore, makeup costs due to losses in the coke filter cake are finally presented as a function of the sulfur removed at the removal efficiency of 95 percent.

At constant energy utilization, the carbonate makeup cost for loss due to the KCl filter cake and the KCl production is as follows:

Cost contribution (3) = (Li_2CO_3 unit cost x lb of Li_2CO_3 makeup per lb of KCl filter cake + K_2CO_3 unit cost x lb of K_2CO_3 makeup per lb of KCl filter cake + Na_2CO_3 unit cost x lb of Na_2CO_3 makeup per lb of KCl filter cake + K_2CO_3 unit cost x lb of K_2CO_3 reacted per lb of KCl produced) x lb/hr of KCl removed at the fly ash filter x 8760 hr/yr x plant load factor

The potassium carbonate reacts with chlorides in the flue gas, such as HCl, which are present due to the chlorine content of the fuel:

$$\underset{138.2\ \text{lb/lb mol}}{K_2CO_3} + \text{chlorides} \longrightarrow \underset{149.1\ \text{lb/lb mol}}{2KCl} + \ldots$$

Thus, the lb of K_2CO_3 reacted per lb of KCl produced is equal to 138.2/149.1 or 0.926, and this additional makeup ratio is shown in Table IV-3 along with the others.

The Kellogg base case operating cost estimate of Table B-6 would have employed a KCl removal rate corresponding to 0.03 percent chlorine content in the coal. This rate was gained proportionately in order to obtain makeup costs corresponding to the percent chlorine content in the coal of 0.01 percent, 0.05 percent, and 0.10 percent. According to Reference 23, chloride content of American coals can range up to 0.5 percent by weight. Thus, makeup costs due to the losses from KCl filter cake and production are presented versus energy utilization and percent chlorine content in the fuel.

The makeup ratios involving Li_2CO_3 incorporate in their magnitudes the basic assumption of 88 percent Li_2CO_3 recovery.

The unit costs employed in the above equations for the lithium, potassium, and sodium carbonates are from Table B-6. These unit costs were compared to those employed in Reference 21 and those quoted recently from two local suppliers. Deviations in unit cost were small for Li_2CO_3 and K_2CO_3. The Kellogg unit cost for K_2CO_3 was approximately one-half of the three others after comparison. Deviations in unit cost were greatest for the Na_2CO_3. However, the Na_2CO_3 makeup cost is the least contributor to the total makeup cost. Therefore, usage of the Kellogg unit costs gives a reasonable representation of the carbonate makeup annual replacement costs.

Other operating costs include utilities, direct labor, supervision, maintenance, overheads, and solids disposal. These are shown as a function of energy utilization and sulfur content of the fuel.

D. Process Developmental Status

1. Classification

One difficulty encountered in the cost evaluation of SO_2 removal processes, is that the processes are at various stages of development. Table IV-4 from Reference 24 classifies most of the known SO_2 removal processes that are under development. The classification criteria at the bottom of the table were established in an American Petroleum Institute study and reported in 1969. The ordering of the processes within a given classification is random.

The dry limestone, wet limestone, and the Monsanto catalytic oxidation processes are classified as "first generation." The magnesium oxide process under development by Chemico has been classified as "advanced second generation." The modified chamber process developed by Tyco has been also classified here because advanced pilot plant studies are presently underway. The molten carbonate process is classified as "second generation" because only bench studies and conceptual testing have been performed thus far by Atomics International. Therefore, cost comparisons between these processes run the risk of being based on nonuniform performance and economic data sources.

TABLE IV-4. DEVELOPMENTAL CLASSIFICATION OF SO_2 REMOVAL PROCESSES

	Company	Process
1.	First Generation	
	Combustion Engineering	Dry Dolomite Injection Wet Scrubbing
	Monsanto	Catalytic Oxidation
	U.S. Bureau of Mines	Alkalyzed Alumina
	Reinluft	Dry Ab-Adsorption
	Several	Dry Dolomite Injection Dry Removal
2.	Advanced Second Generation	
	B&W-Esso	Dry Adsorption & Regeneration
	Mitsubishi	Dap Manganese Oxide
	Wellman Lord	Potassium Sulfite
	Kiyoura	Catalytic Oxidation
	Showa Denko	Ammonia Scrubbing
	Firma Carl Still	Lignite Ash Still
	Lurgi	Wet Char Sulfacid
	Hitachi	Activated Carbon Adsorption
	Ovitron	Wet Scrub-High Acceleration
	Chemico	Magnesium Oxide Scrub
	Tyco	Modified Chamber
3.	Second Generation	
	A. J. Teller	Teller Chromatographic
	J. Sieth and Siemens	Ferrous Oxide
	USBM-Mitsubishi	Red Mud
	USBM	Phosphate Rock
	Chevron Research	Catalytic Reduction
	Princeton Chem Research	Catalytic Reduction
	Ionics Inc.	Alkaline Scrubbing with Electrolytic Cells
	Mitsubishi	Manganese Oxyhydride
	Atomics International	Molten Carbonate
	Consolidation Coal	Potassium Formate Scrub

Classification Criteria

I. First Generation
- a. Advanced Pilot Plant Studies
- b. Active Research and Development
- c. Adaptability to U.S. Market
- d. Available Data on Economic Assessment

II. Advanced Second Generation
- a. Meeting First Generation Criteria Except:
  1. Less Advanced Pilot Studies or Insufficient Information
  2. Better Adaptation to Foreign Operations

III. Second Generation
- a. Essentially Bench Studies of New Concepts
- b. Pilot Studies Incomplete
- c. Economic Assessment Difficult

2. Problem Areas

Some of the problem areas that are unique for each given process are identified as a result of the current course of technological development.

a. Dry Limestone. The dry limestone injection technique has been studied intensely. It exhibits the disadvantage of being able to remove under 50 percent of the sulfur oxides contained in the flue gas. Current TVA test data indicates that at 2.0 stoichiometry, the SO_2 removal efficiency varie between 18 and 26 percent ±12 percent (Ref. 8).

Besides the problem of low removal efficiency, there are several other potential problem areas, none of which have been sufficiently studied to quantify their effects. The addition of limestone into the combustion zone may adversely affect the combustion process, especially at high stoichiometric ratios. Increased slagging caused by the effect of CaO on the fusability of coal ash may inhibit heat transfer. Increased erosion within the boiler may increase maintenance costs. Electrostatic precipitation may not operate as efficiently due to the elimination of SO_3 from the flue gas with subsequent increase in the average particle resistivity. As resistivity is increased, the precipitation rate rate decreases. Also as stated earlier, the addition of limestone base compounds in the ash is likely to increase the problems associated with ash disposal. Different methods of disposal may have to be instituted or existing ones engineered to eliminate the possibility of water pollution.

b. Wet Limestone. Contacting the flue gas with a limestone slurry significantly increases SO_2 removal efficiency. The major problem of the dry limestone process is thus overcome. Efficiency in excess of 85 percent can be achieved today with prospects of improvement to 95 percent and beyond. The problems of boiler erosion and reactant disposal identified in the case of the dry limestone process must also be considered here. However, since the stoichiometric ratios required approach unity, the magnitude of these problems are decreased. Since no electrostatic precipitator is required, the problem of lower precipitator effectiveness is eliminated. The wet limestone process, however, has several unique problem areas which must be investigated. Corrosion and scaling within the scrubber are major problems which must be overcome. Engineers are confident that the study of slurry chemistry with respect to equilibrium composition, pH, degree of supersaturation, rate of desupersaturation, and the role played by precipitated salts will result in satisfactory methods which will control both the corrosion and scaling with the scrubber and associated components.

c. Magnesium Oxide. Several problem areas were discussed in Reference 16 concerned with the pilot demonstration performed by Babcock and Wilcox. Crystalline deposition of magnesium sulfite hydrate in piping, pumps, sumps, and spray nozzles was a recurring problem. It was believed to result from the neutralization reaction between bisulfite and magnesia. Temperature cycling of the scrubbing slurry was also found to contribute to this deposition. The deposition was prevented by the presence of fly ash through a scouring action in the absorption slurry. Nonwettable materials also reduced the tendency to receive deposits. If formed, the deposition can be removed by simple dissolution.

Deposition of solids where the flue gases first contact the spray slurry was a problem which occurred in various degrees for both types of absorbers. These deposits occur near the top of the throat of the venturi absorber. They also formed at the inlet of the floating bed absorber. These deposits were identified as being an aggregate of fly ash and $MgSO_3 \cdot 3H_2O$.

d. Catalytic Oxidation. Problem areas associated with the Cat-Ox process were discussed in Reference 17. Principal problems have to do with fly ash. The fly ash must be removed from the flue gas with very high efficiency in a hot, dry process. Dust not removed from the entering flue gas tends to plug the catalyst bed in the converter, contaminate the product acid, and plug the fiber-bed mist eliminator. Monsanto uses an electrostatic precipitator (or a combination of cyclone separators followed by a precipitator) for the hot, dry gas cleaning operation. The rate at which the catalyst bed plugs depends upon the quantity of dust passing through the precipitator. When the catalyst bed becomes plugged, it is necessary to remove the catalyst for cleaning. The cleaning operation requires a shutdown of two to three days and involves the loss of about 2.5 percent of the catalyst by attrition and breakage. Monsanto estimates that it will be necessary to clean the catalyst at three-month intervals if the dust passing to the catalyst bed does not exceed the quantity allowed for in the design. The required dust collector efficiency for fly ash from firing of pulverized coal is greater than 99.6 percent, which is higher than is now being provided commercially in fly ash precipitators.

Of the dust that passes through the converter, a portion is collected in the acid in the absorption tower and the remainder is collected with very high efficiency by the fiber-bed mist eliminator, where the buildup of solids produces an increase in the resistance to gas flow. Monsanto reports that it has developed a mist eliminator that can be washed periodically, while on stream, to remove the accumulated solids.

The amount of maintenance required for the Cat-Ox system will be acutely dependent on the performance of the electrostatic precipitator. A small deterioration of the precipitator performance may increase the quantity of dust entering the converter by a large factor, with a corresponding reduction in the length of the intervals between shutdowns for cleaning. It will therefore be necessary to attain high standards of maintenance and operation of the precipitators.

Cat-Ox systems are provided in multiple, parallel independent trains, so that part of the system can continue to operate even if one train should fail. The Cat-Ox reheat system can be bypassed in the event of complete failure. If, however, part or all of the integrated Cat-Ox system should fail, it is necessary to reduce the load on the boiler or shut it down. Consequently, the converter in each train is further divided into parallel modular sections that can be dampered off to permit catalyst cleaning if this should be necessary between scheduled shutdowns.

The by-product produced by the Cat-Ox system, 78 percent sulfuric acid contaminated with a small amount (less than 0.1 percent) of fly ash, has a disadvantage from an economic standpoint. Unless it can be transported by barge, its movement to markets is severely limited by the cost of shipment. This is also true, though in lesser degree, of the more concentrated acids, but the Cat-Ox product acid is additionally limited in applications by its lower concentration and its impurities.

e. Modified Chamber. Problem areas pertaining to the Tyco process were discussed in Reference 19. In this system, the scrubber must perform as a particulate removal device for sulfuric acid mist and residual fly ash. It is also an absorber for nitrogen oxides and any unhydrated sulfur trioxide. Major scrubber design and performance parameters need to be confirmed. Additional information is needed on the effects of fly ash on the scrubber system.

Problem areas of the Tyco process include (1) the complexity of process control and operation, (2) the corrosiveness of the scrubbing acid, (3) the contamination of fly ash residue in the major product, sulfuric acid, (4) the need for extremely high efficiency scrubbing to recover nitrogen oxides, and (5) the high system pressure drop required to prevent sulfuric acid mist from escaping into the atmosphere. Process control and operation of the Tyco process will be difficult due to the critical nitrogen dioxide recycle requirements and the low concentrations of nitrosyl-sulfuric acid in the system. Close control will be necessary to prevent excess nitrogen dioxide or nitric oxide losses in the scrubber as changes occur in flue gas compositions. Startups have to be quick and smooth since it is not practical to store or replenish large quantities of nitrogen dioxide for startups.

With the use of boiler feedwater to remove excess heat from the circulating acid and maintain a 250°F temperature, some fluctuation of boiler performance may be encountered during process startup as the switch is made from preheating the acid with steam to cooling the acid with boiler feed waters. Proper instrumentation has been recommended to reduce the problem.

The profitability of the Tyco process will be strongly influenced by the sales value of the by-product acids, 80 percent H_2SO_4 and 60 percent HNO_3. In the operating cost estimates of Appendix B acid unit sales revenues of $4.80/ton and $15/ton, respectively, were employed. The most desirable acids are those with high concentration and purity. Nitric acid produced in the process has these characteristics, but the sulfuric acid has less adequate concentration and purity.

f. Molten Carbonate. Major process and engineering problem areas still requiring solutions prior to full-scale application were identified in Reference 21. One of these was the elimination of melt carryover with the absorber off-gas. This loss of melt is undesirable from an economic standpoint. It is not acceptable from a corrosion standpoint because the scrubbed gases are returned to the boiler where the materials of construction are not likely to be compatible with the melt. Furthermore, the melt carryover may be emitted to the atmosphere from the boiler stack causing additional pollution problems.

In the absorber, good SO_x removal efficiencies occur at high velocities, approximately 25 fps. The optimum liquid-to-gas ratio (total mols of melt/ total mols of gas) is low at 0.0138 and may not be achievable. The freezing point of the melt is also not too far below the nominal absorber operating temperature of 850°F.

The sequence of reduction reactions and the required internal design of the reducer need to be demonstrated. The effect of the elevated operating temperature of the reducer, 1500°F, on materials of construction is required. The design of a vessel with three dissimilar materials in contact with each other presents design problems because of differential expansion. The two-zone reducer concept is another area which requires additional study to insure operability. The mode for introducing the coke, e.g., the addition of coke as a slurry with the melt, is another area where further test work is required.

A primary problem exists to find fly ash and coke filters capable of producing a solids cake of low liquid content at the process temperature of 850°F. The "dry cake" discharges carry measurable quantities of melt. These melt quantities must be minimized by means of a satisfactory dry cake discharge unit in order to reduce the cost for molten carbonate replacement.

Further studies are required to determine the extent to which the carbonate components can be economically recovered from the solids filter cakes. A recovery process has been assumed for the lithium carbonate. It has been assumed that the potassium and sodium salts are not recovered after the aqueous lithium carbonate recovery process because:

(1) The fly ash filter cake contains the KCl which is soluble along with the other potassium and sodium salts.

(2) The coke filter cake contains the heavy metals which are removed along with the soluble potassium and sodium salts.

The discarded sodium and potassium salts are mixtures of carbonates, sulfites, sulfates, and sulfides in addition to the potassium chloride. The heavy metals are nickel and vanadium contained in the original coke.

E. Desulfurization of Fuels

An alternate method of SO_2 abatement is to desulfurize the fuel before it is burned. Both oil and coal can be desulfurized. This section deals with the economics involved in oil desulfurization, both domestic and Caribbean, as well as with the mechanical cleaning of coal. Mechanical cleaning refers to the removal of pyrites from the coal which tends to reduce the total sulfur content of the coal. Coal gasification or other of the more complex coal desulfurization methods were not studied since these possibilities will not be commercially possible until after the 1970s.

1. Residual Fuel Oil Desulfurization

The economics of residual fuel oil (RFO) desulfurization lack the precision that has been achieved with respect to the normal oil processing costs. In part, this is due to the absence of long-term operating data. Large-scale desulfurization has emerged as a feasible operation only in recent years. The prediction of incremental costs due to desulfurization processing is complicated by the large number of independent variables. For example, RFOs of similar sulfur contents which are desulfurized to the same levels can exhibit significant differences in processing costs. Reactor size will vary drastically from resid to resid. Thus, capital investment and catalyst costs can differ significantly. The rate of hydrogen consumption differs depending on process parameters and extent of desulfurization. Metal content in the resid determines catalyst life. Generally, specific desulfurization cost estimates must be based on the characteristics of the feed stock, on the level of desulfurization require, refinery size, and on regional factors such as construction, labor, utility, and hydrogen costs.

With respect to appreciation of desulfurization cost with time, it is believed that the savings due to process improvements will outweigh the increased costs of labor and materials for some time to come. Catalysts with ever-increasing resistance to poisoning by sulfur, nitrogen, metals, coke-forming materials, and other bad actors in RFO are being found (Ref. 25). It is expected that the cost of desulfurization will have a downward trend in the foreseeable future assuming the present trend in technological improvements continues.

One of the most extensive efforts in the evaluation of RFO desulfurization costs to date was undertaken by Bechtel Corporation for the American Petroleum Institute (API) (Ref. 26). The study analyzed 14 desulfurization schemes with respect to a typical Caribbean fuel oil normally containing 2.6 weight-percent sulfur. The usefulness of the results obtained in this study derives from the fact that the bulk of the RFO imported into the United States comes from the Caribbean refineries. For example, in 1966, about 3.4 million barrels from an import total of 3.8 million barrels were Caribbean RFO.

The various constraints and considerations adopted by Bechtel in evaluating incremental desulfurization costs are summarized below:

(a) Initial sulfur content of 2.6 weight-percent
(b) Typical large Caribbean refinery size of 300,000 barrels per stream day of crude oil
(c) A 57.4 volume percent yield of desulfurized No. 6 residual fuel oil
(d) Initial metal content of 500 ppm
(e) Costs derived from incremental investment and operating costs; any taxes or transportation costs were not included
(f) Costs include sulfur credit based on a market price of $32 per long ton; sulfur credit had only a minor effect on the overall incremental costs (maximum credit was about 11 percent of final incremental cost)

In order to provide a means of evaluating desulfurization costs for Caribbean oils of slightly different sulfur contents, some extrapolation was applied to the Bechtel cost estimates. Thus, there are three curves corresponding to imported Caribbean oils in the nomograph. However, care should be taken not to venture too far from the 2.6 weight-percent sulfur baseline. Desulfurization costs are highly sensitive to the type of processing facilities and the overall properties of the oil feed stock. Whenever a process involves hydrogen, for example, a major cost item is the hydrogen feed. This cost can vary drastically with facility location.

The cost differentials shown in the nomograph are based on the assumption that all of the residual oil must be desulfurized. It may be economically advantageous to desulfurize only part of the resid. One option presented in

Reference 27 is to desulfurize three-quarters of the resid and sell the other quarter, which is high in sulfur, to users which have stack gas controls. Under this scheme, the incremental cost of the desulfurized oil would be 50 percent of that shown on the nomograph. Another plan would concentrate the sulfur into a very high sulfur coke which would be used by the refinery for process heat. With the proper incentives, the refinery could install an efficient SO_2 removal system. The overall economics of this alternative appears promising for those refineries which are "energy poor," having no immediate source of inexpensive fuel. If such alternatives are instituted, then the cost of desulfurization per barrel obtained from the nomograph must be considered as being conservatively high.

To provide flexibility in treating capital finance, the nomographs allow variation in the fixed charge rate. Figure B of Nomograph page 1 may be used to determine the capital recover factor. Add to this the recurring costs, such as insurance and taxes, which results in the fixed charge rate.

The Bechtel Corporation assumed that money would be available at six percent interest with a payout time of five years. Allowing for about two percent of recurring costs, a fixed charge rate of 25 percent results. Thus, a typical refinery would fix its prices to recover one-quarter of its capital investment every year.

The fixed and variable costs of desulfurization are calculated separately in the nomograph. Variable and fixed costs must be added to obtain the total operating cost. For instance, if we require that the Caribbean oil (average sulfur content = 2.6 weight-percent) be reduced to 0.5 pounds of sulfur per million Btu's, the following costs are calculated:

Capital investment	44¢/10^6 Btu/yr
Fixed cost (FCR-25%)	11¢/10^6 Btu
Variable cost	3.4¢/10^6 Btu
Total cost	14.4¢/10^6 Btu

This is equivalent to about 90 cents per barrel of oil having a heating value of 6.3 x 10^6 Btu per barrel.

Incremental desulfurization costs for domestic RFO are based on results of a study performed by Arthur G. McKee and Company (Ref. 28). In this study, incremental costs were evaluated for sulfur levels of 1.0 weight-percent and 0.5 weight-percent, according to the five individual Petroleum Administration for Defense (PAD) districts.* The basic approach in the study was to establish an "average" refinery within each PAD district. Each average refinery was characterized with respect to crude properties, product rate, product characteristics, processing schemes, and RFO blending requirements.

A major assumption was made with respect to the blending of the heavy residuum with desulfurized cutter stock. It was assumed that the desulfurized cutter stock was already available as part of the overall effort in maintaining a constant sulfur level in the overall domestic RFO pool. Thus, the incremental costs reflect only the processing necessary for desulfurizing the heavy residuum.

In the case of PAD Districts 2 and 3, for example, the lower sulfur content of the crudes obviates additional desulfurization processing. The 1.0 weight-percent sulfur level in these districts can be achieved simply by blending with low sulfur cutter stock. Hence, the incremental desulfurization cost is zero. To obtain the 0.5 weight-percent sulfur level, it is necessary to partially desulfurize the heavy residuum in all five PAD districts.

Table IV-5 summarizes the McKee data. Note that the processing scheme used produced two products: No. 2 and No. 6 fuel oils. This complicates the determination of desulfurization costs. If one assumes that there is a strong market for the additional No. 2 fuel oil, a credit can be applied to the costs since No. 2 oil may run as high as $9 per barrel as compared with No. 6 oil at $2 per barrel. However, any real credit is highly dependent on local market conditions. Under ideal conditions, where No. 2 oil can be sold at a high price, the cost of desulfurization (allowing a No. 2 oil credit) can be reduced 50 percent or more depending on the PAD district. In constructing the nomographs, no credit was given for the No. 2 oil. This conservative assumption is partially counteracted by the fact that the McKee study assumed the cutter stock was made available for blending at no cost. What, in effect, has been done is to assume that No. 2 and No. 6 oils are essentially the same and that all the costing is done on a per-barrel-of-new-product basis.

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*The Department of the Interior continues to employ this identification of different geographical areas in the United States. The districts are: I- East, II - Midwest, III - Southern, IV - Rocky Mountain, V - West.

TABLE IV-5. DOMESTIC DESULFURIZATION DATA FOR UNITED STATES PAD DISTRICTS
(Based on Arthur G. McKee and Company Data)

	DISTRICT				
	1	2	3	4	5
Minimum Refinery Capacity (b/d)	70,000	40,000	40,000	10,000	42,500
Crude Characteristics					
API Gravity	29.7	35.1	34.0	32.6	27.1
Sulfur (wt %)	1.33	0.73	0.73	1.39	1.29
Resid Data					
Sulfur Without Cutter (wt %)	2.6	1.3	1.3	2.7	2.5
Sulfur With Cutter (wt %)	2.0	1.1	1.1	2.25	1.9
Old No. 6 Production (b/d)	11,340	4,640	5,600	2,810	10,631
1% Sulfur Level					
New No. 6 Production (b/d)	9,350	---	---	2,250	15,470
No. 2 Production (b/d)	2,570	---	---	730	4,100
Total New Products (b/d)	11,920	---	---	2,980	19,570
Incremental Investment (10^6\$)	7.43	---	---	3.19	10.49
Variable Cost (\$/day)	4,372	---	---	2,625	5,627
Capital Investment (¢/10^6Btu/yr)*	28.8	---	---	49.5	24.7
Variable Cost (¢/10^6Btu)*	5.8	---	---	14.0	4.6
0.5% Sulfur Level					
New No. 6 Production (b/d)	7,940	3,710	4,480	1,910	13,220
No. 2 Production (b/d)	4,310	1,210	1,455	1,152	6,890
Total New Product (b/d)	12,250	4,920	5,935	3,062	20,110
Incremental Investment (10^6\$)	8.84	5.19	5.72	3.82	11.38
Variable Cost (\$/day)	5,527	3,565	3,858	3,435	6,507
Capital Investment (¢/10^6Btu/yr)*	33.4	48.8	44.6	57.7	26.2
Variable Cost (¢/10^6Btu)*	7.2	11.5	10.3	17.8	5.1

*See text for definitions and assumptions.

The nomographs plot capital investment in terms of cents per million Btu per year and variable cost in terms of cents per million Btu. These parameters were obtained from the basic data as follows:

$$\text{Capital investment } (¢/10^6 \text{ Btu/yr}) = \frac{(100)\ II}{(6.3)(365)TNP \times PF}$$

$$\text{Variable cost } (¢/10^6 \text{ Btu}) = \frac{(100)\ VC}{(6.3)\ TNP}$$

where:

II = incremental investment (dollars)

TNP = total new product (barrels per day)

PF = plant factor, assumed to be 0.94

VC = variable cost (dollars per day)

100 = cents in a dollar

6.3 = million Btu's in a barrel

365 = days in a year

The fixed and variable costs for desulfurization of domestic oils are nomographed. Variable and fixed costs must be added to obtain the total incremental cost for desulfurization. For instance, in PAD I if we require a limit of 0.5 pounds of sulfur per million Btu to be maintained, the following costs are calculated:

Capital investment	32.0 $¢/10^6$ Btu/yr
Fixed cost (FCR-25%)	8.0$¢/10^6$ Btu
Variable cost	6.7$¢/10^6$ Btu
Total cost	14.7$¢/10^6$ Btu

This is equivalent to about 93 cents per barrel of oil having a heating value of 6.3 x 10^6 Btu per barrel.

Most desulfurization processes,including the ones used in the Bechtel and McKee studies, rely on either thermal or catalytic treatment of RFO. This leads to separation of lighter end products from the residual oil. The result is a net loss in the oil's heating value. A typical heating value for No. 6 oil is 6.3 x 10^6 Btu per barrel. Desulfurization to low sulfur values (~0.5 percent sulfur) can lower the heating value to about 5 x 10^6 Btu per barrel. This effect is included in the nomograph.

2. Coal Desulfurization

a. Introduction. Sulfur is found in raw coal in two principal forms, organic and pyritic (FeS_2) sulfur. Free sulfur exists in coal, but only in minute quantities. Sulfate sulfur is also present. It varies considerably due to weathering of coal which oxidizes small amounts of pyrite. This sulfur in sulfates normally is considered as part of the organic sulfur in the data.

Sulfur can be removed in many ways. The organic sulfur is chemically bound in a complex manner to the hydrocarbon compounds. Its removal requires drastic treatments, many of which also result in removal of the pyritic sulfur. Solvent extraction, liquefication, and gasification have all been studied as possible methods of removing some, if not all, of the sulfur. Gasification appears to offer the best chance for development. However, a commercially competitive process has yet to be discovered. The advent of improved gasifiers and the continuing increase in the prices of the established fuels could alter this situation.

Pyrite occurs in coal in bands, in pockets, or as particles, fine or coarse, mixed with the coal. Pyrite in banks or pockets or as coarse particles is relatively easy to separate by known washing techniques. Very fine pyrite becomes increasingly difficult to separate as its size is reduced below that of the crushed coal size. Before the bulk of the pyrite can be removed from the coal, the coal must first be crushed fine enough to release the pyrites. Once released, the pyrite may be removed by electrostatic, magnetic, froth flotation, concentrating tables, or bacterial methods. The principal method now being investigated is the use of concentrating tables. Such tables, which use the difference in specific gravity of the pyrite to separate it, can remove as much as 70 percent of the pyrite sulfur (Ref. 29).

The present study was limited to the economic analysis of conventional coal cleaning methods which utilize concentrating tables for removing the bulk of the pyritic sulfur.

b. Coal Cleaning Potential. The removal of sulfur from coal by purely physical means is an inherently limited method of desulfurization. Removing

70 percent of the pyrite usually results in only a moderate decrease in the total sulfur content. The extent of total reduction depends largely on the proportional amounts of pyritic and organic sulfur present in the coal. Typically, total sulfur reductions of between 20 and 40 percent can be achieved. However, it is difficult to generalize because of the considerable variations in the pyrite-organic sulfur ratio and in the washability characteristics of specific coals.

It is known that a limited number of U.S. steam coals can be desulfurized in an economic fashion by pyrite removal. The question may be asked whether these coals make up a worthwhile portion of the steam coal market. Figure IV-13 shows the coal reserves of the United States in terms of coal type and sulfur content. It should be noted that these data are subject to question since typifying a coal field's average sulfur content based on several sulfur measurements can be meaningless because of the extreme variability in sulfur within any one given seam. In any case, the data shown are considered a best estimate. Henceforth, the discussion will be limited to bituminous coal. This coal type makes up the majority of the coal for steam production.

When bituminous coal is pulverized and subjected to a washing process, certain amounts of the pyritic sulfur and ash are removed. The degree of removal is dependent, in part, upon the specific gravity of separation. This parameter is varied by the design and operation of the concentration tables. As the specific gravity of separation is decreased, more pyrite and ash are removed. Figure IV-14 shows the relation between specific gravity and removal. Note that a substantially higher percentage of ash is removed than sulfur. It must be kept in mind that the removal of ash represents, in itself, an important improvement. A reduction in ash results in a saving in coal transportation, in boiler maintenance, in dust removal equipment, and in ash disposal. Thus, when studying the economics of desulfurization, a credit for ash removal must be considered which defrays part of the process costs. On the other hand, as specific gravity of separation is decreased, an increasing percentage of the combustible material is also removed along with the ash and sulfur. This is reflected in the "percent recovery" curve in Figure IV-14. In economic terms, the loss of heating value associated with the combustible material removed is considered the overriding cost variable which works against the feasibility of further removal of sulfur or ash. The cost implications of ash removal and heating value loss will be discussed in the next section.

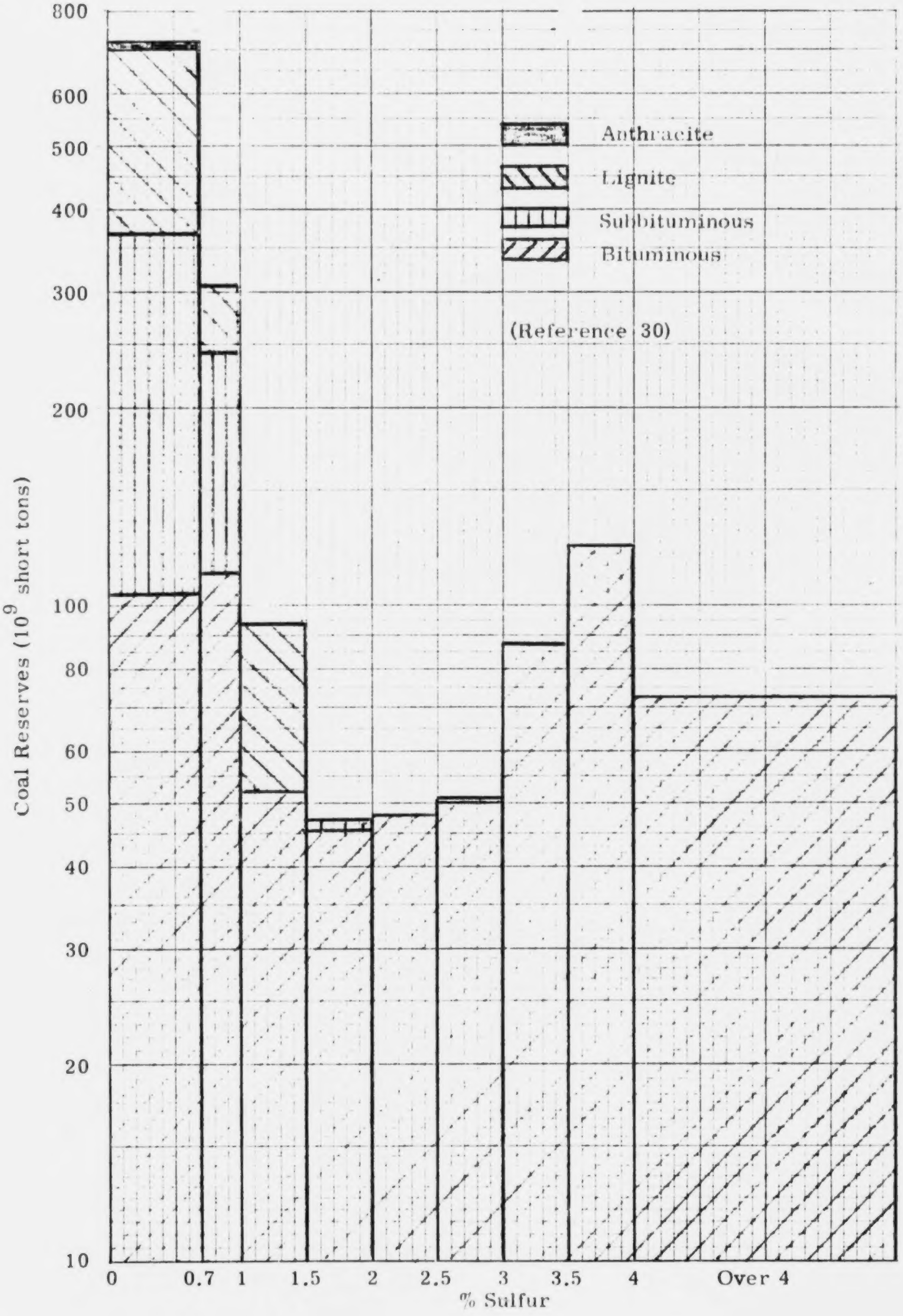
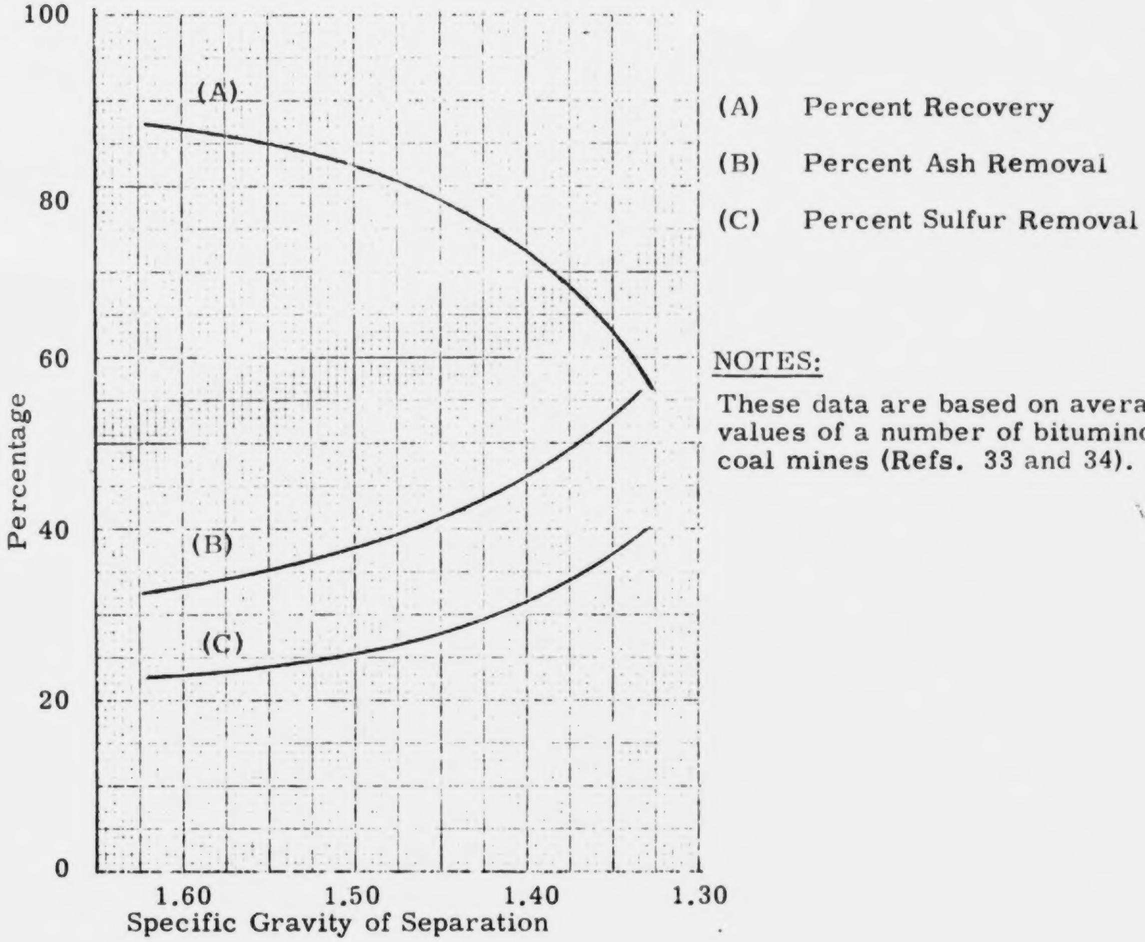


Figure IV-13. Estimated Remaining Coal Reserves of the United States on January 1, 1965



NOTES:

These data are based on average values of a number of bituminous coal mines (Refs. 33 and 34).

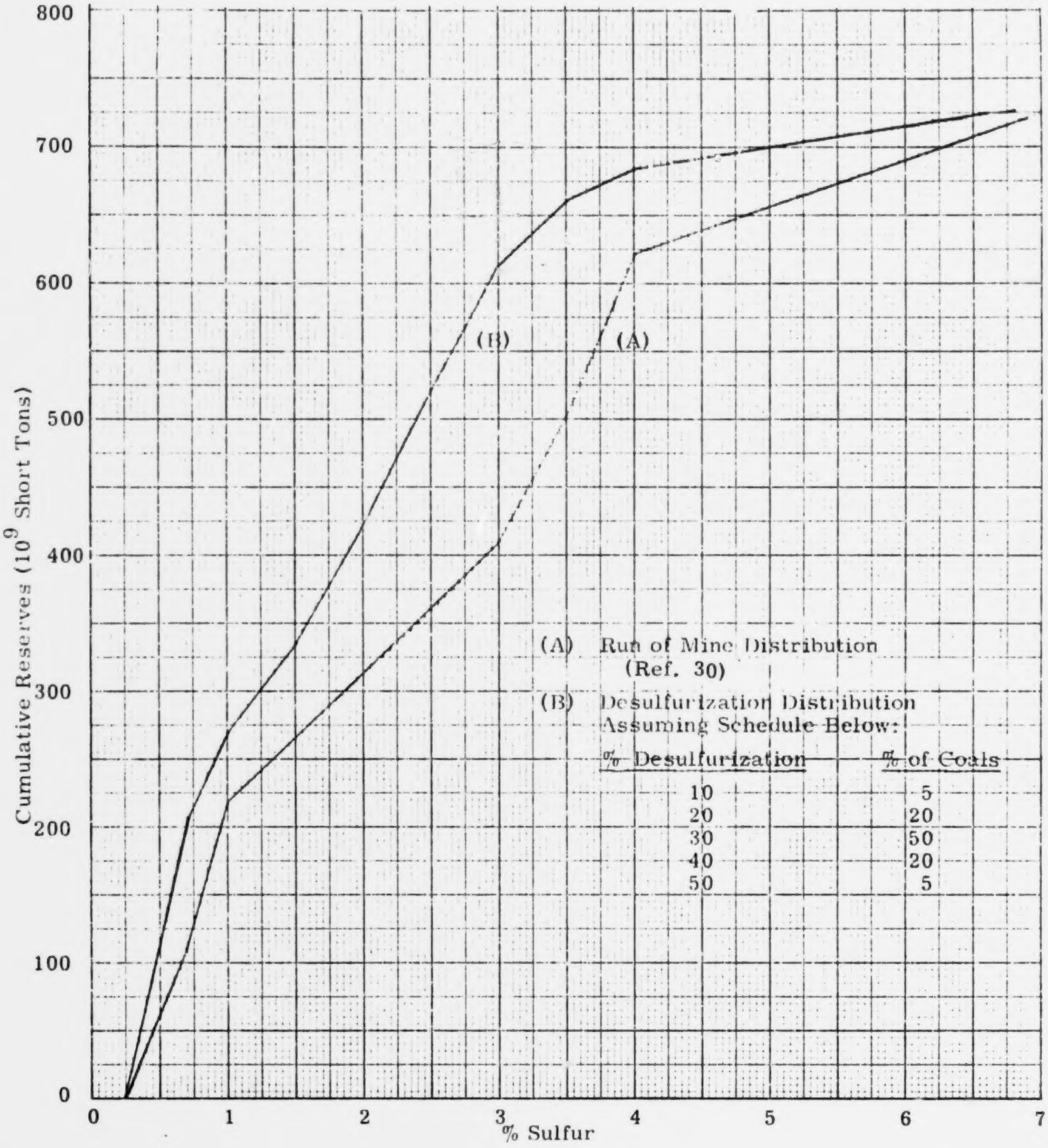
Figure IV-14. Effect of Specific Gravity of Separation on Coal Washing Parameters

The potential of pyrite removal by coal cleaning is dependent upon the extent of desulfurization possible in the coal fields of interest. References 31 through 35 present data and discussions related to desulfurization potential. Emphasis is placed on eastern bituminous coal. Figure IV-14 has summarized these data in terms of averages. Based on a review of the data, the following desulfurization schedule was proposed for bituminous coals:

Percent Desulfurization	Applicable % of Coals
0	0
10	5
20	20
30	50
40	20
50	5
>50	0

The schedule assumes that a certain percentage of the United States bituminous coals can be economically desulfurized by the percentages given. This schedule was used to estimate the coal cleaning potential of the known bituminous coals. Figure IV-15 shows the results. Curve A is the cumulative reserve having a sulfur content of a given percentage of sulfur or less. Curve B is the coal's sulfur distribution assuming all the coal is cleaned according to the assumed schedule. The curves may be interpreted as follows: If a one percent sulfur limit were placed on coal-fired utilities, then there exist some 216×10^9 short tons of bituminous coal which may be burned with no additional desulfurization equipment or coal washing. If the coals having somewhat higher sulfur contents were processed, then 50×10^9 additional short tons of coal would become available (the difference between curves B and A). Thus, on a reserve basis, there would exist some 23 percent more coal than otherwise would be available at one percent or less. This is the case for the U.S. as a whole. Of course, if put on an individual coal region or local market basis, the results would, in some cases, be more promising. Thus, coal cleaning as a significant and economical general alternative for SO_2 abatement is probably limited to specific regions or mines.

c. Coal Desulfurization Economics. A coal processing plant must be designed for a specific coal. No two coals are identical and thus no one system



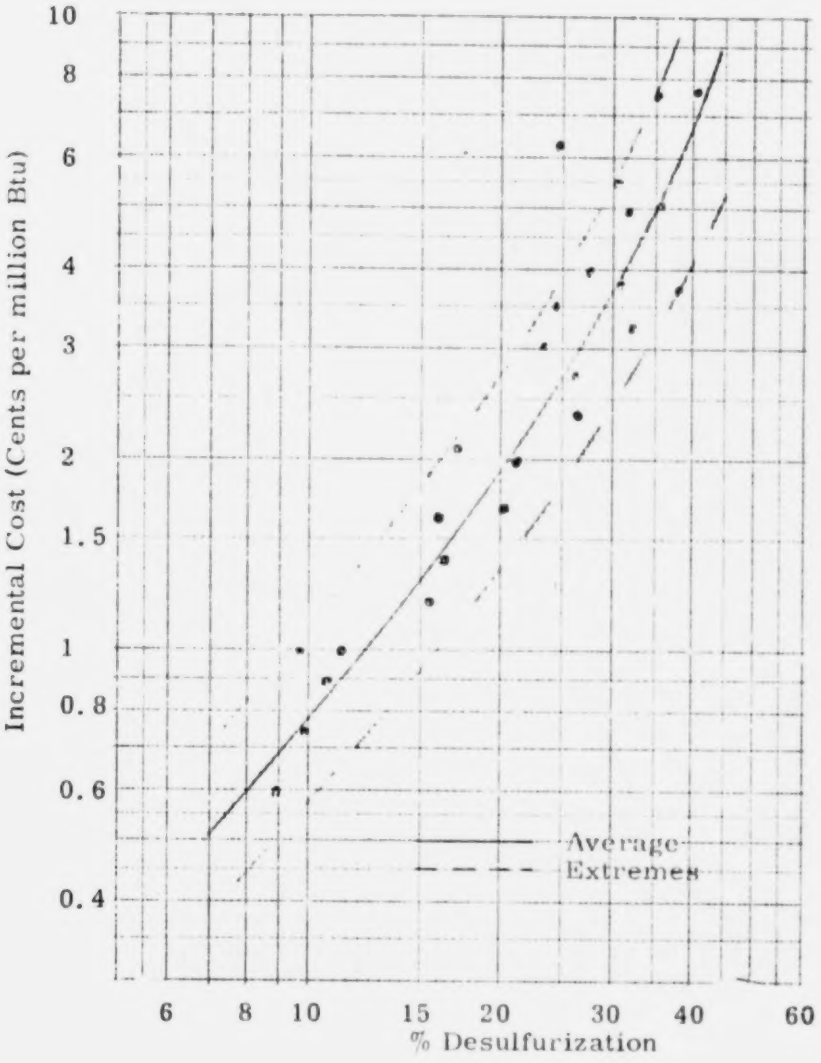
% Desulfurization	% of Coals
10	5
20	20
30	50
40	20
50	5

Figure IV-15. Sulfur Distribution of Remaining Bituminous Coal Reserves (1965)

can do an optimum job of desulfurizing them all. The properties of coal coming from the same seam may vary to such an extent that system efficiency must be compromised to accept the variability. Sulfur content varies as does the form in which it is present. Pyrite liberation and washability of the coal varies. Thus, each installation must be tailor-made (Ref. 37). This factor introduces scatter in the cost data as do mining costs and variation in coal sizes. This must be kept in mind when typical cost data are presented.

The cost of coal processing was based on two independent studies, the Paul Weir Company study in 1965 and the Resources of the Future, Inc., study in 1968 (Refs. 31 and 32). The general approach of these studies was to take specific eastern bituminous mines, study their existing mining and process methods, and then to upgrade the processing to implement different degrees of desulfurization. Normally, three levels of desulfurization were investigated; i.e., those which correspond to separation densities of 1.60, 1.45, and 1.35. Process complexity and cost increase with each step.

Because of the variability in the data, a number of different forms of data presentation were studied in an effort to minimize cost variability. The form which appeared to minimize variability made cost dependent upon the percent of desulfurization. Figure IV-16 depicts this relationship based on the Paul Weir 1965 costs. Considering the vast differences in coal and plant characteristics, the resulting scatter in cost is reasonably small. Though not presented, the Resources of the Future, Inc., 1968 cost data for 13 eastern mines generally fell within the dashed lines. A comparison of the ground rules utilized in the two studies was made. It was concluded that the Weir study resulted in somewhat higher costs. Thus, the solid line shown in Figure IV-16 was selected as the "typical" cost relation (with the knowledge that it was a slightly conservative average of the available data). This line was then appreciated based on the Marshall and Stevens Mining Industry Index (Ref. 36). An appreciation factor of 1.4 was used to convert 1965 dollars to 1971 dollars. This appreciation is applicable to both the capital and operating costs. The fixed and variable costs were then separated and a nomograph constructed. This nomograph shows capital investment and variable cost as a function of the run-of-the-mine (ROM) sulfur content and the ultimate sulfur content required. For example, assume that a coal has a four percent sulfur content



Costs Based on a typical plant size of 850 tons/hr and plant amortized over 20 years (Zero interest rate)

Figure IV-16. Incremental Desulfurization Costs Based On Five Reference Bituminous Coal Mines (1965 Dollars)

required and it is desulfurized 30 percent. Thus, the ROM content is four percent and the ultimate content is 2.8 percent. The capital investment is then 9¢/10^6 Btu/yr. Using a fixed charge rate of ten percent per year, a fixed cost of 0.90¢/10^6 Btu results. The variable cost is found to be 4.4¢/10^6 Btu. The total operating cost is 5.3¢/10^6 Btu. This is equivalent to $1.27/ton for a 12,000 Btu/lb final product.

The above costs do not account for any credits which might be obtained from the increased heating value and lower ash content of the coal. In all cases, there is a saving in terms of ash disposal and maintenance costs associated with the ash. Reference 38 analyzed the cost of removal of ash from power plant stacks and ash disposal. It was determined that present rates are about 10¢/ton of coal burned. Reference 39 gives TVA maintenance costs which are associated with the presence of ash. Average maintenance costs of about 10¢/ton of coal burned were reported. Since coal processing typically removes from 25 to 50 percent of the original ash, a credit of 5-10¢/ton of coal results. Several desulfurization processes keep the as-washed coal size large enough to allow shipment by train. Based on coal transportation costs and the known average mileage between mine and utility, a typical transport cost of $3/ton was assumed. Since the average Btu value of the processed coal is about five percent greater than unprocessed coal, a saving of about 15¢/ton can be realized. Thus, looking at the extremes, a total credit due to improvements in the coal lie between 5 and 25¢/ton of coal burned. This is approximately 0.2 to 1.0¢/10^6 Btu. It can be seen that the credits obtained from the improved coal partially negate the fixed cost of the process. Thus, one can calculate the desulfurization costs considering only the variable costs with little loss in accuracy.

F. Transportation Costs

Some abatement methods require a change in the pattern of fuel transportation or require the shipment of bulk reactants. In an attempt to account for the added costs resulting from changes or additions in material transportation, a nomograph has been constructed which allows one to estimate transportation costs.

An example of one possible use of the transportation nomograph is to analyze the trade-off between using a stack gas SO_2 control versus western low sulfur coal. The incremental cost of the stack control process in terms of mills/Kw-hr can be compared with the premium cost of low sulfur western coal. The differential costs of the coals at the mine mouth plus the differential rail costs can be computed resulting in an incremental cost convertible to mills/Kw-hr. Thus, an economic comparison can be made which would identify the least expensive control method.

The transportation rates used to construct the cost curves are based on 1968 to 1970 data. They are "average" rates for this time period. As more efficient methods of transportation begin to dominate a certain transport mode (e.g., unit trains or super tankers), the average rate should be reduced. Also, the cost of certain transport modes can vary from day to day or from region to region because of local factors such as the supply and demand of that particular mode of transportation. Long-term contracts and special agreements for shipping large quantities can also result in costs different from the industry's "average." These market problems are discussed for the more important modes in the text which follows.

1. Oil Transportation Costs

The cost of transporting oil involves more variation of costs than any other fuel. Costs may range from one cent per barrel per 100 miles for long haul tankers to 45 cents per barrel per 100 miles for short hauls by rail.

a. Oil Transport Costs Via Tanker. The tanker market is a highly volatile and competitive market. Tonnage is contracted on a voyage basis ("spot" market) or on short-term or long-term charters. The spot market can fluctuate from day to day over a wide range. If spot prices are kept high for several weeks, they may have a decided inflationary impact on the charter prices contracted during the period of high spot prices. Because rates fluctuate so erratically, it was decided to present what is termed the "base rates" for tanker transport on the nomograph. The current rate structure can then be factored in using the "World Scale" factor.

Base rates have been placed on all the important runs; e.g., Caribbean-United States, Persian Gulf-U.K., or Persian Gulf-Far East runs.

These base rates are equated to a World Scale of 100 (W100). If rates for a given run happen to be twice the base rate, it would be termed as a rate of W200.

As an example, the Caribbean-United States run is 2000 miles. The World Scale base rate at W100 is 25 cents per barrel. In the last half of 1970, spot charter rates varied from W180 to over W320 with an average near W240 (Ref. 40). Current World Scale rates appear weekly in the Oil and Gas Journal.

Another important example is the Persian Gulf-U.K. run which is 11,300 miles. This run has significance because it largely influences the rates of other runs. Its base rate at W100 is $1.19 per barrel. Actual rates have varied between W120 and W300 with an average of over W200 for the last part of 1970. These rates are very high and are deemed unusual for the industry. In coming years, with the influx of the super tankers into the fleet and the possible advent of the Suez Canal opening, demand should be lessened and rates brought back to more "normal" levels. Under normal market conditions, the super tankers in the 300,000 dwt class would probably operate the Persian Gulf-U.K. run at below W50 which, no doubt, will substantially affect the total market price structure (Ref. 41).

b. Oil Transport Via Rail. The cost of rail transportation has generally decreased since the beginning of the 1960s. Though it has been uncommon to ship crude or residual oils in tanker cars in the past, this mode of transportation has been on the increase. Canada has recently begun the first unit train operation transporting well head oil from its northern provinces. Such unit trains drastically cut the cost, operating below one-half the normal 25 cents per barrel per 100 mile rate reported for long haul operation greater than 500 miles. Unlike the ocean tanker market, the rail market is less volatile, being dependent on more long-term influences. Variations in rates are due mainly to regional cost differences in such items as labor costs, level of modernization, and local tariff structure. The rates used to construct the nomograph represent average rates. Rates can vary ±20 percent. Unit trains are excluded. Rates for this mode would be approximately one-half the indicated cost per barrel.

c. Oil Transport Via Barge. Barge rates level off beyond 500 miles. The rates for these long hauls average about 16 cents per barrel per 100 miles. Significant cost reductions are foreseen in this area as barges carrying 80 to 100 thousand barrels come into use. Long-term charters of such large barges will significantly affect the cost structure bringing the cost of oil transportation down to levels as low as five cents per barrel per 100 miles (Ref. 42). Care should be exercised in using the average barge rate shown on the nomograph because, as in the case of ocean tankers, rates are highly variable.

d. Oil Transport Via Pipeline. Pipeline transportation costs are presently averaging about five cents per barrel per 100 miles for long distance pipelines which operate over 8000 hours per year at capacity (Ref. 42). Costs skyrocket for short pipelines which have low utilization. As an example, a 7.6 mile pipeline built by Consumers Power between Edmonton and Bay City, Michigan, costs the company 46 cents per barrel (Ref. 43). This is equivalent to a rate of $6.00 per barrel per 100 miles, which is about 120 times more costly than the basic long distance rate. Thus, the nomograph which is based on high utilization pipelines should not be used for short-run "connect" pipelines which are subject to considerable variability in cost due to their intermittent operation.

2. Coal and Bulk Solids Transportation Costs

a. Bulk Solids Transport Via Truck. Several of the SO_2 removal processes require a supply of rather expensive reactants or catalysts. This material will probably be brought by truck. Thus, for convenience in calculating delivered reactant cost, typical truck shipping rates have been nomographed. These assume a fully-loaded van carrying a minimum shipment of 37,000 pounds (Ref. 10).

b. Coal Transport Via Rail. Coal shipped by rail has normally cost about one to three cents per ton-mile. The introduction of long haul unit trains will reduce this rate to about one-half cent per ton-mile. Even lower rates are projected for utilities requiring large supplies of coal. Such utilities are planning modern loading-unloading terminals to reduce turnaround time and will purchase their own unit trains. The nomograph includes two curves, one for single loads and one for unit train operation (Refs. 10 and 42).

c. Coal Shipments Via Barge. For relatively long hauls (~300 miles), the TVA estimates barge rates for coal between two and three mills per ton-mile depending on the annual tonnage of coal contracted. The nomograph presentation has assumed an annual tonnage of several million tons per year. Adding a $0.17 per ton handling cost brings the long haul rate of this annual quantity to about three mills per ton-mile (Ref. 10).

3. Gas Transportation

Only one method of transporting gas across the country, pipeline, is presently used. Long distance rates for pipelines of greater than 500 miles run between 0.8 and 1.5 cents per thousand cubic feet per 100 miles (Ref. 42). The transportation of liquified natural gas by tanker from international sources is not yet a significant factor in the U.S. fuel supply picture.

V. CONCLUSIONS AND RECOMMENDATIONS

The major conclusions and recommendations resulting from the contract activities are summarized below:

(1) SO_2 process economics can be nomographed with reasonable accuracy. In most cases, a nomograph of a process can be limited to three or four basic cost elements. Secondary economic variables such as labor rates and utility rates can be assumed constant with little effect on overall accuracy.

(2) The parametric study which was performed using the process nomographs exemplify the flexibility and usefulness of the nomographic method. They show that the nomographs can predict control costs for nearly any set of input parameters.

(3) Several nomograph formats were considered during the course of the contract. The format which was initially selected placed an entire set of curves on a single foldout. This format resulted in a compact group of nomographs which could be utilized conveniently with a limited amount of page turning. An alternate format which was finally chosen placed curves on a single page. This method though requiring more page manipulations is desirable for nomograph applications since it provides curves of greater accuracy and allows for duplication of curves of special interest.

(4) Nearly all cost estimates utilized in the study were based on dissimilar ground rules. Data presentation formats differed. Assumptions as to what costs were and were not chargeable to the abatement process differed. The methodology of charging peripheral cost elements such as contractor fees, indirect field costs, and contingency varied significantly. Some estimates were complete while others gave little backup information. Thus, it is strongly recommended, based on the difficulties faced in evaluating current economic data,

that a uniform cost estimate guideline be prepared specifically for SO_2 stack gas processes. Such a guideline should include a uniform format for cost presentation. It should define what is and what is not chargeable to the process. Also, it should prescribe what data must be submitted as backup to the cost estimate.

(5) The nomographs as presented in this report are based on the best currently available economic data. It is expected, however, that these data will have to be modified as process development continues. Prototype tests currently being conducted or in the planning stages are expected to make significant contributions to economic data base. Improvements in, or modifications to, system components are also expected to affect economics. Thus, it is recommended that the data be reviewed periodically and the nomographs updated when necessary. Also, those systems which have not been nomographed but which are likely to reach commercialization should be studied and nomographs constructed. This would include the Chemico version of the magnesium oxide process which differs substantially from the Babcock and Wilcox process which was analyzed in the current study.

(6) Current EPA activities involving the development and demonstration of SO_2 control processes should be reviewed with the objective of assuring that any economic data generated by their activities be made available. Emphasis should be placed on those cost elements which are not well known (such as maintenance and wear-out of components, waste disposal, or additive consumption).

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APPENDIX A

NOMOGRAPH EXAMPLES

(Refer to respective nomographs)

TABLE A-1. PLANT SIZE AND FUEL CONSUMPTION NOMOGRAPH SAMPLE CASE

A.	Heat Rate		
	Plant efficiency	(enter)	38
	Heat rate, 10^3 Btu/Kw-hr	(read)	~9000
B.	Capital Recovery Factor		
	Interest rate, %	(enter)	8
	Years amortized	(enter)	15
	Capital recovery factor	(read)	11.5
C.	Energy Utilization		
	Net generating capacity, Mw	(enter)	500
	Plant load factor, %	(enter)	75
	Annual energy production, 10^6 Kw-hr/yr	(read)	3.25
	Plant heat rate, Btu/Kw-hr	(enter)	9000
	Energy utilization, 10^{12} Btu/yr	(read)	29.6
D.	Fuel Cost and Consumption		
	Energy utilization, 10^{12} Btu/yr	(enter)	29.6
	Heating value of oil	(enter)	----
	Annual fuel consumption, 10^7 bbl/yr	(read)	0.46
	Fuel cost, $/bbl	(enter)	3.5
	Energy cost, ¢/10^6 Btu	(read)	55.0
E.	Sulfur Content Unit Conversion		
	Weight-percent of sulfur in fuel	(enter)	3.5
	Heating value, oil	(enter)	----
	Sulfur content, lb/10^6 Btu	(read)	1.75
F.	Flue Gas Flow Rate		
	Net generating capacity, Mw	(enter)	500
	Plant heat rate, Btu/Kw-hr	(enter)	9000
	Flow rate, 10^6 scfm	(read)	0.9

TABLE A-2. DRY LIMESTONE NOMOGRAPH SAMPLE CASE

Net generating capacity, Mw	200
Plant heat rate, Btu/Kw-hr	1,100
Plant load factor, %	50
As-received heating value of coal, Btu/lb	12,000
Weight-percent of sulfur in fuel	2.0
Sulfur content of fuel, lb/10^6 Btu	1.67
Energy utilization, 10^{12} Btu/yr	9.6
Flow rate, 10^6 scfm	0.44

A. Efficiency

Stoichiometric ratio	(enter)	2.5
SO_2 removal efficiency (optimistic), %	(read)	30

B. Limestone Use Factor

Sulfur content, lb/10^6 Btu	(enter)	1.67
Stoichiometric ratio	(enter)	2.5
Percent purity	(enter)	90
Limestone use factor	(read)	14.3

C. Fixed Cost

Flow rate, 10^6 scfm	(enter)	0.44
Limestone use factor	(enter)	14.3
Capital cost, \$$10^6$	(read)	1.15
Fixed charge rate, %/yr	(enter)	20
Fixed cost, \$$10^6$/yr	(read)	0.23

D. Other Costs

Energy utilization, 10^{12} Btu/yr	(enter)	9.6
Limestone use factor, lb/10^6 Btu	(enter)	14.3
Other costs, 10^6 \$/yr	(read)	0.35

E. Fine Grinding Correction

Limestone particle size, microns	(enter)	-----
Incremental cost of grinding	(read)	0

F. Additive cost

Energy utilization, 10^{12} Btu/yr	(enter)	9.6
Limestone use factor, lb/10^6 Btu	(enter)	14.3
Limestone cost, \$/ton		4.0
Incremental grinding cost, \$/ton		0
Disposal cost, \$/ton		0.5
Total reactant cost, \$/ton	(enter)	4.5
Annual additive cost, \$/yr	(read)	0.32

$\therefore$ Total Operating Cost $= C + D + E + F$

$= 0.23 + 0.35 + 0.32 + 0$

$= 0.90 \times \$10^6/\text{yr}$

TABLE A-3. WET LIMESTONE NOMOGRAPH SAMPLE CASE

Net generating capacity, Mw	500
Plant heat rate, Btu/Kw-hr	9,000
Plant load factor, %	75
As-received heating value of coal, Btu/lb	12,000
Weight-percent of sulfur in fuel	3.5
Sulfur content of fuel, lb/10^6 Btu	2.92
Energy utilization, 10^{12} Btu/yr	29.6
Flow rate, 10^6 scfm	0.9

A. Blank (an overall SO_2 removal efficiency of 85% is assumed)

B. Limestone Use Factor

Sulfur content, lb/10^6 Btu	(enter)	2.92
Stoichiometric ratio	(enter)	1.30
Percent purity	(enter)	90
Limestone use factor, lb/10^6 Btu	(read)	13.5

C. Fixed Cost

Flow rate, 10^6 scfm	(enter)	0.9
Limestone use factor	(enter)	13.5
Capital cost, $\$10^6$	(read)	5.8
Fixed charge rate, %/yr	(enter)	20
Fixed cost, $\$10^6$/yr	(read)	1.16

D. Other Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Limestone use factor, lb/10^6 Btu	(enter)	13.5
Other costs, $\$10^6$/yr	(read)	1.22

E. Reheat Correction

Flow rate, 10^6 scfm	(enter)	----
Reheat temperature, °F	(enter)	----
Credit, $\$10^6$/yr	(read)	0

F. Additive Cost

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Limestone use factor, lb/10^6 Btu	(enter)	13.5
Limestone cost, \$/ton		4.0
Disposal cost, \$/ton		0.50
Total reactant cost, \$/ton	(enter)	4.50
Annual additive cost, \$/yr	(read)	0.90

$$
\begin{aligned}
\therefore \text{ Total Operating Cost } &= C + D - E + F \\
&= 1.16 + 1.22 - 0 + 0.90 \\
&= 3.28 \times \$10^6/\text{yr}
\end{aligned}
$$

TABLE A-4. MAGNESIUM OXIDE NOMOGRAPH SAMPLE CASE

Net generating capacity, Mw	500
Plant heat rate, Btu/Kw-hr	9,000
Plant load factor, %	75
As-received heating value of coal, Btu/lb	12,000
Wight-percent of sulfur in fuel	3.5
Sulfur content of fuel, lb/10^6 Btu	2.92
Energy utilization, 10^{12} Btu/yr	29.6
Flow rate, 10^6 scfm	0.9

A. Fixed Cost

Flow rate, 10^6 scfm	(enter)	0.9
Sulfur content, lb/10^6 Btu	(enter)	2.92
Fixed capital cost, \$$10^6$	(read)	16.35
Fixed capital cost, \$$10^6$	(enter)	16.35
Fixed charge rate, %/yr	(enter)	20
Fixed cost, \$$10^6$/yr	(read)	3.27

B. By-Product Acid Credit

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Sulfur content, lb/10^6 Btu	(enter)	2.92
Sulfur removed, 10^3 tons/yr	(read)	43.0
Sulfur removed, 10^3 tons/yr	(enter)	43.0
98.5% H_2SO_4 quantity, 10^3 tons/yr	(read)	130.
98.5% H_2SO_4 unit cost, \$/ton	(enter)	10
98.5% H_2SO_4 credit, \$$10^6$/yr	(read)	1.30

C. Raw Materials Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Magnesia quantity, 10^3 tons/yr	(read)	0.655
Magnesia unit cost, \$/ton	(enter)	40
Magnesia cost, \$$10^6$/yr	(read)	0.0275

D. Other Operating Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Sulfur content, lb/10^6 Btu	(enter)	2.92
Other operating costs, \$$10^6$/yr	(read)	2.24

$\therefore$ Total Operating Cost $= A - B + C + D$

$= 3.27 - 1.30 + 0.0275 + 2.24$

$= 4.24 \times \$10^6/\text{yr}$

TABLE A-5. CATALYTIC OXIDATION NOMOGRAPH SAMPLE CASE

Net generating capacity, Mw	500
Plant heat rate, Btu/Kw-hr	9,000
Plant load factor, %	75
As-received heating value of coal, Btu/lb	12,000
Weight-percent of sulfur in fuel	3.5
Sulfur content of fuel, lb/10^6 Btu	2.92
Energy utilization, 10^{12} Btu/yr	29.6
Flow rate, 10^6 scfm	0.9

A. Fixed Cost

Flow rate, 10^6 scfm	(enter)	0.9
Sulfur content, lb/10^6 Btu	(enter)	2.92
Fixed capital cost, $\$10^6$	(read)	25.1
Fixed capital cost, $\$10^6$	(enter)	25.1
Fixed charge rate, %/yr	(enter)	20
Fixed cost, $\$10^3$/yr	(read)	5.02

B. By-Product Acid Credit

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Sulfur content, lb/10^6 Btu	(enter)	2.92
Sulfur removed, 10^3 tons/yr	(read)	40.0
Sulfur removed, 10^3 tons/yr	(enter)	40.0
78% H_2SO_4 quantity, 10^3 tons/yr	(read)	150
78% H_2SO_4 unit cost, $/ton	(enter)	5
78% H_2SO_4 credit, $\$10^6$/yr	(read)	0.75

C. Raw Materials Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Ash content in fuel, wt-%	(enter)	10
V_2O_5 catalyst quantity, ft^3/yr	(read)	425.0
V_2O_5 catalyst quantity, ft^3/yr	(enter)	425.0
V_2O_5 unit cost, $/$ft^3$	(enter)	40
V_2O_5 cost, $\$10^6$/yr	(read)	0.017

D. Other Operating Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Other operating costs, $\$10^6$/yr	(read)	2.32

$\therefore$ Total Operating Cost = A - B + C + D

= 5.2 - 0.75 + 0.017 + 2.32

= 6.79 x $\$10^6$/yr

TABLE A-6. MODIFIED CHAMBER NOMOGRAPH SAMPLE CASE

Net generating capacity, Mw	500
Plant heat rate, Btu/Kw-hr	9,000
Plant load factor, %	75
As-received heating value of coal, Btu/lb	12,000
Weight-percent of sulfur in fuel	3.5
Sulfur content of fuel, lb/10^6 Btu	2.92
Energy utilization, 10^{12} Btu/yr	29.6
Flow rate, 10^6 scfm	0.9

A. Fixed Cost

Flow rate, 10^6 scfm	(enter)	0.9
Fixed capital cost, \$$10^6$	(read)	18.0
Fixed capital cost, \$$10^6$	(enter)	18.0
Fixed charge rate, %/yr	(enter)	20
Fixed cost, \$$10^6$/yr	(read)	3.6

B. Raw Materials Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Granular charcoal unit cost, $/ft	(enter)	15
Catalyst cost, \$$10^6$/yr	(read)	0.475

C. Other Operating Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Other operating costs, \$$10^6$/yr	(read)	2.2

D. By-Product Acid Credits

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Sulfur content of fuel, lb/10^6 Btu	(enter)	2.92
80% H_2SO_4 quantity, 10^3 tons/yr	(read)	142.0
Sulfur removed, 10^3 tons/yr	(read)	39.9
80% H_2SO_4 quantity, 10^3 tons/yr	(enter)	142.0
80% H_2SO_4 unit cost, $/ton	(enter)	5
80% H_2SO_4 credit, \$$10^6$/year	(read)	0.71
Energy utilization, 10^{12} Btu/yr	(enter)	29.6
NO_x volume content in flue gas, ppm	(enter)	400
60% HNO_3 quantity, 10^3 tons/yr	(read)	16.2
60% HNO_3 quantity, 10^3 tons/yr	(enter)	16.2
60% HNO_3 unit cost, $/ton	(enter)	15
60% HNO_3 credit, \$$10^6$/yr	(read)	.243

∴ Total Operating Cost = A + B + C - D

= 3.6 + 0.475 + 2.2 - (0.71 + .243)

= 5.33 x \$$10^6$/yr

TABLE A-7. MOLTEN CARBONATE NOMOGRAPH SAMPLE CASE

Net generating capacity, Mw	500
Plant heat rate, Btu/Kw-hr	9,000
Plant load factor, %	75
As-received heating value of coal, Btu/lb	12,000
Weigh-percent of sulfur in fuel	3.5
Sulfur content of fuel, lb/10^6 Btu	2.92
Energy utilization, 10^{12} Btu/yr	29.6
Flow rate, 10^6 scfm	0.9

A. Fixed Cost

Flow rate, 10^6 scfm	(enter)	0.9
Sulfur content, lb/10^6 Btu	(enter)	2.92
Fixed capital cost, $\$10^6$	(read)	22.2
Fixed capital cost, $\$10^6$	(enter)	22.2
Fixed charge rate, %/yr	(enter)	20
Fixed cost, $\$10^6$/yr	(read)	4.44

B. Other Operating Costs

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Sulfur content, lb/10^6 Btu	(enter)	2.92
Other operating costs, $\$10^6$/yr	(read)	2.46

C. By-Product Sulfur Credit

Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Sulfur content, lb/10^6 Btu	(enter)	2.92
Sulfur removed ≈ sulfur by-product, 10^3 tons/yr	(read)	42.0
Sulfur removed ≈ sulfur by-product, 10^3 tons/yr	(enter)	42.0
Sulfur unit cost, \$/ton	(enter)	10
Sulfur credit, $\$10^6$/yr	(read)	0.42

D. Raw Materials Costs

Sulfur removed, 10^3 tons/yr	(enter)	42
Coke utilization, 10^3 tons/yr	(read)	58
Coke unit cost, \$/ton	(enter)	10
Coke cost, $\$10^6$/yr	(read)	0.58
Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Ash content in fuel, wt-%	(enter)	10
M_2CO_3 makeup cost (fly ash filter), $\$10^6$/yr	(read)	0.065
Sulfur removed, 10^3 tons/yr	(enter)	42
M_2CO_3 makeup cost (coke filter), $\$10^6$/yr	(read)	0.18
Energy utilization, 10^{12} Btu/yr	(enter)	29.6
Chlorine content in fuel, wt-%	(enter)	0.05
M_2CO_3 makeup cost (KCl filter), $\$10^6$/yr	(read)	.240

∴ Total Operating Cost = A + B - C + D
= 4.44 + 2.46 - 0.42 + 0.240 + 0.58 + 0.065 + 0.18
= 7.55 x $\$10^6$/yr

APPENDIX B

COST ESTIMATES

TABLE B-1. ANNUAL OPERATING COSTS DRY LIMESTONE PROCESS

NOTE: M = 10^3
MM = 10^6

Plant size, Mw	350
Fixed Capital Investment (FCI)	$3,400,000
Stream time, hr/yr	7000

			$/yr
DIRECT COST			
1.	Operating labor (1 man/shift @ $4.50/hr)		$ 39,420
2.	Supervision - 15% of Item 1		5,910
3.	Maintenance labor and materials - 4% of FCI		136,000
4.	Plant supplies - 15% of Item 3		20,400
5.	Utilities		
	a. Cooling water	- --- GPM at $0.025/M gal	----
	b. Process water	- --- GPM at $0.20/M gal	----
	c. Electric power	- 4420 Kw at 6.75 mills/Kw-hr	208,850
	d. Natural gas	- --- MSCFH at $0.40/MSCF	----
	e. Steam	- --- M lb/hr at $0.45/M lb	----
6.	Chemicals and raw materials		
	a. Limestone	- 716 TPD @ $2/ton	417,700
	b. Extra coal*	- 1.05 TPH @ $8/ton	58,800
6a.	Solids disposal (ex fly ash)	- 29 TPH @ $0.50/ton	101,500
7.	Subtotal Direct Cost (ex credits)		$ 988,580
8.	Credits		
	None		----
9.	Total Direct Cost		$ 988,580
INDIRECT COST			
10.	Payroll overhead	- 20% of (1 + 2)	9,070
11.	Plant overhead	- 50% of (1 + 2 + 3 + 4)	100,870
12.	Total Indirect Cost		$ 109,940
FIXED COST			
13.	Capital charges (includes depreciation, interim replacements, insurance, taxes, and cost of capital)	- 18.22% of FCI	$ 619,480
TOTAL OPERATING COST			
14.	Net Production Cost --	Items (9 + 12 + 13)	$1,718,000
UNIT PRODUCTION COST			
15.	Gross	- Items (7 + 12 + 13)	$1,718,000
	a. mills/Kw-hr		0.70
	b. $/ton S not emitted		112.2
16.	Net	- Items (9 + 12 + 13)	$1,718,000
	a. mills/Kw-hr		0.70
	b. $/ton S not emitted		112.2

*Extra coal needed for limestone calcination in boiler

TABLE B-2. ANNUAL OPERATING COSTS WET LIMESTONE PROCESS

Note: M = 10^3
MM = 10^6

Plant size, Mw	1000
Fixed capital investment	$10,800,000
Stream time, hr/yr	7000

	DIRECT COST			$/yr
1.	Operating labor (2 men/shift @ $4.50/hr)			78,840
2.	Supervision - 15% of Item 1			11,830
3.	Maintenance labor and materials - 4% of FCI			432,000
4.	Plant supplies - 15% of Item 3			64,800
5.	Utilities			
	a. Cooling water	-	--- GPM at $0.025/M gal	----
	b. Process water	-	--- GPM at $0.20/M gal	----
	c. Electric power	-	25,310 Kw at 6.75 mills/Kw-hr	1,195,860
	d. Steam	-	--- M lb/hr at $0.45/M lb	----
6.	Chemicals and raw materials			
	a. Limestone	-	1237 TPD @ $2/ton	721,700
	b. Extra coal*	-	3 TPH @ $8/ton	168,000
6a.	Solids disposal (ex fly ash)	-	57 TPH @ $0.50/ton	199,500
7.	Subtotal Direct Cost (ex credits)			$2,872,530
8.	Credits			
	None			----
9.	Total Direct Cost			$2,872,530
	INDIRECT COST			
10.	Payroll overhead	-	20% of (1 + 2)	18,130
11.	Plant overhead	-	50% of (1 + 2 + 3 + 4)	293,740
12.	Total Indirect Cost			$ 301,870
	FIXED COST			
13.	Capital charges - 18.22% of FCI (includes depreciation, interim replacements, insurance, taxes, and cost of capital)			$1,967,750
	TOTAL OPERATING COST			
14.	Net production cost	-	Items (9 + 12 + 13)	$5,142,160
	UNIT PRODUCTION COST			
15.	Gross	-	Items (7 + 12 + 13)	$5,142,160
	a. mills/Kw-hr			0.73
	b. $/ton S not emitted			65.2
16.	Net	-	Items (9 + 12 + 13)	$5,142,160
	a. mills/Kw-hr			0.73
	b. $/ton S not emitted			65.2

*Extra coal needed for limestone calcination in boiler

TABLE B-3. ANNUAL OPERATING COSTS MAGNESIUM OXIDE PROCESS

Note: M = 10^3
MM = 10^6

Plant size, Mw	1000
Fixed capital investment (FCI) -	\$26,800,000
Stream time, hr/yr	7000

				$/yr
DIRECT COST				
1.	Operating labor (4 men/shift @\$4.50/hr)			\$ 157,680
2.	Supervision - 15% of Item 1			23,650
3.	Maintenance labor and materials - 4% of FCI			1,072,000
4.	Plant supplies - 15% of Item 3			160,800
5.	Utilities			
	a.	Cooling water -	4230 gpm at \$0.025/M gal	44,420
	b.	Process water -	616 gpm at \$0.20/M gal	51,740
	c.	Electric power -	22,656 Kw at 6.75 mills/Kw-hr	1,070,500
	d.	Natural gas -	268 MSCFH at \$0.40/MSCF	750,400
	e.	Steam -	M lb/hr at \$0.45/M lb	---
6.	Chemicals and raw materials			
	a.	Magnesia -	5 TPD @ \$40/ton	58,330
	b.	Petroleum coke -	645 lb/hr (dry) @ \$8/ton	18,060
	c.	Limestone -	1500 lb/hr @ \$2/ton	10,500
7.	Subtotal Direct Cost (ex credits)			\$3,418,080
8.	Credits			
	a.	Sulfuric acid (98.5%) -	890 TPD @ \$12/ton	(3,115,000)
9.	Total Direct Cost			\$ 303,080
INDIRECT COST				
10.	Payroll overhead	-	20% of (1 + 2)	36,270
11.	Plant overhead	-	50% of (1 + 2 + 3 + 4)	707,070
12.	Total Indirect Cost			\$ 743,340
FIXED COST				
13.	Capital charges (Includes depreciation, interim replacements, insurance, taxes, and cost of capital)	-	18.22% of FCI	\$4,882,960
TOTAL OPERATING COST				
14.	Net Production Cost	-	Items (9 + 12 + 13)	\$5,929,380
UNIT PRODUCTION COST				
15.	Gross	-	Items (7 + 12 + 13)	\$9,044,380
	a. mills/Kw-hr			1.29
	b. \$/ton S not emitted			111.1
16.	Net	-	Items (9 + 12 + 13)	5,929,380
	a. mills/Kw-hr			0.85
	b. \$/ton S not emitted			71.3

TABLE B-4. ANNUAL OPERATING COSTS CAT-OX PROCESS

NOTE: M = 10^3
MM = 10^6

Plant size, Mw	1000
Fixed Capital Investment (FCI)	$43,400,000
Stream time, hr/yr	7000

DIRECT COST

			$/yr
1.	Operating labor (4 men/shift @ $4.50/hr)		$ 157,680
2.	Supervision - 15% of Item 1		23,650
3.	Maintenance labor and materials - 4.5% of FCI		1,953,000
4.	Plant supplies - 15% of Item 3		292,950
5.	Utilities		
	a. Cooling water -	548 GPM at $0.025/M gal	5,760
	b. Process water -	--- GPM at $0.20/M gal	----
	c. Electric power -	17,800 Kw at 6.75 mills/Kw-hr	841,050
	d. Natural gas -	--- MSCFH at $0.40/MSCF	----
	e. Steam -	--- M lb/hr at $0.45/M lb	----
6.	Chemicals and raw materials		
	a. Sulfuric acid catalyst -	1330 ft^3/yr @ $38.23/$ft^3$	50,850
7.	Subtotal Direct Cost (ex credits)		$ 3,324,940
8.	Credits		
	a. Sulfuric acid (78%) -	1061 TPD @ $6/ton 100%	(1,447,250)
	b. Stack gas cooling* -	162 MM Btu/hr @ $0.40/MM Btu/hr	(453,600)
9.	Total Direct Cost		$ 1,424,070

INDIRECT COST

10.	Payroll overhead -	20% of (1 + 2)	36,270
11.	Plant overhead -	50% of (1 + 2 + 3 + 4)	1,213,640
12.	Total Indirect Cost		$ 1,249,910

FIXED COST

13.	Capital charges - (includes depreciation, interim replacements, insurance, taxes, and cost of capital)	18.22% of FCI	$ 7,907,480

TOTAL OPERATING COST

14.	Net Production Cost -	Items (9 + 12 + 13)	$10,581,460

UNIT PRODUCTION COST

15.	Gross -	Items (7 + 12 + 13)	$12,482,330
16.	a. mills/Kw-hr		1.78
	b. $/ton S not emitted		158.0
16.	Net -	Items (9 + 12 + 13)	10,581,460
	a. mills/Kw-hr		1.51
	b. $/ton S not emitted		133.9

*Credit for stack gas at 225°F vs base case 300°F

TABLE B-5. ANNUAL OPERATING COSTS MODIFIED CHAMBER PROCESS

NOTE: M=10^3
MM=10^6

Plant size, Mw	1000
Fixed Capital Investment (FCI)	$29,800,000
Stream time, hr/yr	7000

			$/yr
DIRECT COST			
1.	Operating labor (5 men/shift @ $4.50/hr)		$ 197,100
2.	Supervision - 15% of Item 1		29,570
3.	Maintenance labor and materials - 5% of FCI		1,490,000
4.	Plant supplies - 15% of Item 3		223,500
5.	Utilities		
	a. Cooling water	- 1720 GPM at $0.025/M gal	18,060
	b. Process water	- 8 GPM at $0.20/M gal	1,340
	c. Electric power	- 24,730 Kw at 6.75 mills/Kw-hr	1,168,480
	d. Natural gas	- --- MSCFH at $0.40/MSCF	----
	e. Steam	- --- M lb/hr at $0.45/M lb	----
6.	Chemicals and raw materials		
	a. Activated carbon*	- 68,000 ft^3/yr @ $13.50/$ft^3$	918,000
	b. Limestone	- 1.25 TPH @ $2/ton	17,500
6a.	Solids disposal (ex fly ash)	- 1 TPH @ $0.50/ton	3,500
7.	Subtotal Direct Cost (ex credits)		$ 4,067,050
8.	Credits		
	a. Sulfuric acid (80%)	- 971 TPD @ $6/ton 100%	(1,359,500)
	b. Nitric acid (60%)	- 148 TPD @ $25/ton 100%	(648,000)
	c. BFW heating	- 170 MM Btu/hr @ $0.40/MM Btu	(474,900)
9.	Total Direct Cost		$ 1,584,650
INDIRECT COST			
10.	Payroll overhead	- 20% of (1 + 2)	45,330
11.	Plant overhead	- 50% of (1 + 2 + 3 + 4)	970,090
12.	Total indirect cost		1,015,420
FIXED COST			
13.	Capital charges (includes depreciation, interim replacements, insurance, taxes, and cost of capital)	- 18.22% of FCI	$ 5,429,560
TOTAL OPERATING COST			
14.	Net Production Cost	- Items (9 + 12 + 13)	$ 8,029,630
UNIT PRODUCTION COST			
15.	Gross	- Items (7 + 12 + 13)	$10,512,030
	a. mills/Kw-hr		1.50
	b. $/ton S not emitted		137.7
16.	Net	- Items (9 + 12 + 13)	$ 8,029,630
	a. mills/Kw-hr		1.15
	b. $/ton S not emitted		104.7

*Assume one year activated carbon life in catalytic stripper

TABLE B-6. ANNUAL OPERATING COSTS
MOLTEN CARBONATE PROCESS

Note: M = 10^3
MM = 10^6

Plant size, Mw	1000
Fixed Capital Investment (FCI)	$34,600,000
Stream time, hr/yr	7000

			$/yr
	DIRECT COST		
1.	Operating labor (5 men/shift @ $4.50/hr		$ 197,100
2.	Supervision - 15% of Item 1		29,570
3.	Maintenance labor and materials - 5% of FCI		1,730,000
4.	Plant supplies - 15% of Item 3		259,500
5.	Utilities		
	a. Cooling water	- 650 GPM at $0.025/M gal	6,830
	b. Process water	- --- GPM at $0.20/M gal	----
	c. Electric power	- 14,575 Kw at 6.75 mills/Kw-hr	688,670
	d. Natural gas	- --- MSCFH at $0.40/MSCF	----
	e. Steam	- ---- M lb/hr at $0.45/M lb	----
6.	Chemicals and raw materials		
	a. Lithium carbonate	- 115 lb/hr @ $0.42/lb	338,100
	b. Sodium carbonate	- 990 lb/hr @ $0.02/lb	138,600
	c. Potassium carbonate	- 1684 lb/hr @ $0.05/lb	589,400
	d. Petroleum coke	- 15.75 TPH @ $8/ton	882,000
6a.	Solids disposal (ex fly ash)	- 1 TPH @ $0.50/ton	3,500
7.	Subtotal Direct Cost (ex credits)		$ 4,863,270
8.	Credits		
	a. Sulfur	- 261 TPD @ $20/short ton	(1,522,500)
9.	Total Direct Cost		$ 3,340,770
	INDIRECT COST		
10.	Payroll overhead	- 20% of (1 + 2)	45,330
11.	Plant overhead	- 50% of (1 + 2 + 3 + 4)	1,108,090
12.	Total indirect cost		$ 1,153,420
	FIXED COST		
13.	Capital charges (includes depreciation, interim replacements, insurance, taxes, and cost of capital)	- 18.22% of FCI	$ 6,304,120
	TOTAL OPERATING COST		
14.	Net Production Cost	- Items (9 + 12 + 13)	$10,798,310
	UNIT PRODUCTION COST		
15.	Gross	- Items (7 + 12 + 13)	$12,320,910
	a. mills/Kw-hr		1.76
	b. $/ton S not emitted		148.4
16.	Net	- Items (9 + 12 + 13)	$10,798,310
	a. mills/Kw-hr		1.54
	b. $/ton S not emitted		130.1

TABLE B-7. PARAMETRIC STUDY
DRY LIMESTONE PROCESS

PLANT SIZE	350 MW 2% S	350 MW 3.5% S*	350 MW 5% S	175 MW 3.5% S
Percent of Sulfur Removed from Flue Gas	50	50	50	50
Fixed Capital Investment ($MM)	3.2	3.4	3.7	2.3
Salable By-Product	-	-	-	-
By-Product Production (Tons/Day)	-	-	-	-
By-Product Credit ($MM/Yr)	-	-	-	-
Total Gross Production Cost ($MM/Yr)	1.36	1.72	2.09	1.05
$/Ton Sulfur Not Emitted	155.8	112.2	95.5	137.0
Mills/KWH	0.56	0.70	0.85	0.86
% Increase of Power Cost	8.3	10.4	12.6	12.7
Total Net Production Cost ($MM/Yr)	-	-	-	-
$/Ton Sulfur Not Emitted	-	-	-	-
Mills/KWH	-	-	-	-
% Increase of Power Cost	-	-	-	-

*BASE CASE

Notes:

By-Product Values

78-80% Sulfuric Acid	-	$6/Ton 100% Contained
98.5% Sulfuric Acid	-	$12/Ton
60% Nitric Acid	-	$25/Ton 100%
Sulfur	-	$20/Short Ton

Plant Load Factor - 80% (7,000 Hr. Stream Year)
Electric Power Manufacturing Cost - 6.75 Mills/KWH
Plant Location - Cincinnati, Ohio

TABLE B-8. PARAMETRIC STUDY WET LIMESTONE PROCESS

PLANT SIZE	1,000 MW 2% S	1,000 MW 3.5% S*	1,000 MW 5% S	700 MW 3.5% S	400 MW 3.5% S
Percent of Sulfur Removed from Flue Gas	82.5	90	93	90	90
Fixed Capital Investment ($MM)	9.6	10.8	12.0	8.0	4.7
Salable By-Product	-	-	-	-	-
By-Product Production (Tons/Day)	-	-	-	-	-
By-Product Credit ($MM/Yr)	-	-	-	-	-
Total Gross Production Cost ($MM/Yr)	4.01	5.14	6.00	3.76	2.25
$/Ton Sulfur Not Emitted	97.1	65.2	51.4	68.2	71.3
Mills/KWH	0.57	0.73	0.36	0.77	0.80
% Increase of Power Cost	8.4	10.8	12.7	11.4	11.9
Total Net Production Cost ($MM/Yr)	-	-	-	-	-
$/Ton Sulfur Not Emitted	-	-	-	-	-
Mills/KWH	-	-	-	-	-
% Increase of Power Cost	-	-	-	-	-

*BASE CASE

Notes:

By-Product Values
78-80% Sulfuric Acid - $6/Ton 100% Contained
98.5% Sulfuric Acid - $12/Ton
60% Nitric Acid - $25/Ton 100%
Sulfur - $20/Short Ton

Plant Load Factor - 80% (7,000 Hr. Stream Year)
Electric Power Manufacturing Cost - 6.75 Mills/KWH
Plant Location - Cincinnati, Ohio

TABLE B-9. PARAMETRIC STUDY
MAGNESIUM OXIDE PROCESS

PLANT SIZE	1,000 MW 2% S	1,000 MW 3.5% S*	1,000 MW 5% S	700 MW 3.5% S	400 MW 3.5% S
Percent of Sulfur Removed from Flue Gas	91	95	96.5	95	95
Fixed Capital Investment ($MM)	22.3	26.8	31.9	20.5	14.1
Salable By-Product	98.5% H_2SO_4				
By-Product Production (Tons/Day)	487	890	1,289	623	356
By-Product Credit ($MM/Yr)	1.70	3.12	4.51	2.18	1.25
Total Gross Production Cost ($MM/Yr)	7.86	9.04	10.38	6.86	4.57
$/Ton Sulfur Not Emitted	172.3	111.1	86	117.9	137.6
Mills/KWH	1.12	1.29	1.48	1.40	1.63
% Increase of Power Cost	16.6	19.1	21.9	20.7	24.1
Total Net Production Cost ($MM/Yr)	6.16	5.93	5.86	4.68	3.33
$/Ton Sulfur Not Emitted	134.9	71.3	48.6	80.4	100.1
Mills/KWH	0.88	0.85	0.84	0.96	1.19
% Increase of Power Cost	13.0	12.6	12.4	14.2	17.6

*BASE CASE

Notes:

By-Product Values
- 78-80% Sulfuric Acid - $6/Ton 100% Contained
- 98.5% Sulfuric Acid - $12/Ton
- 60% Nitric Acid - $25/Ton 100%
- Sulfur - $20/Short Ton

Plant Load Factor - 80% (7,000 Hr. Stream Year)
Electric Power Manufacturing Cost - 6.75 Mills/KWH
Plant Location - Cincinnati, Ohio

TABLE B-10. PARAMETRIC STUDY CAT-OX PROCESS

PLANT SIZE	1,000 MW 2% S	1,000 MW 3.5% S*	1,000 MW 5% S	700 MW 3.5% S	400 MW 3.5% S
Percent of Sulfur Removed from Flue Gas	82.5	90	93	90	90
Fixed Capital Investment ($MM)	42.3	43.4	44.1	33.2	20.6
Salable By-Product	78% H_2SO_4				
By-Product Production (Tons/Day)	565	1,061	1,552	743	424
By-Product Credit ($MM/Yr)	0.77	1.45	2.12	1.01	0.58
Total Gross Production Cost ($MM/Yr)	12.12	12.48	12.67	9.56	5.94
$/Ton Sulfur Not Emitted	295.5	158.0	109.0	173.5	188.7
Mills/KWH	1.74	1.78	1.81	1.95	2.12
% Increase of Power Cost	25.8	26.4	26.8	28.9	31.4
Total Net Production Cost ($MM/Yr)	10.97	10.58	10.10	8.23	5.18
$/Ton Sulfur Not Emitted	265.8	133.9	86.9	149.3	164.5
Mills/KWH	1.57	1.51	1.44	1.68	1.85
% Increase of Power Cost	23.3	22.4	21.3	24.9	27.4

*BASE CASE

Notes:

By-Product Values

78-80% Sulfuric Acid	-	$6/Ton 100% Contained
98.5% Sulfuric Acid	-	$12/Ton
60% Nitric Acid	-	$25/Ton 100%
Sulfur	-	$20/Short Ton

Plant Load Factor - 80% (7,000 Hr. Stream Year)
Electric Power Manufacturing Cost - 6.75 Mills/KWH
Plant Location - Cincinnati, Ohio

TABLE B-11. PARAMETRIC STUDY
MODIFIED CHAMBER PROCESS

PLANT SIZE		1,000 MW 2% S	1,000 MW 3.5% S*	1,000 MW 5% S	700 MW 3.5% S	400 MW 3.5% S
Percent of Sulfur Removed from Flue Gas		78	87.5	91	87.5	87.5
Fixed Capital Investment ($MM)		29.7	29.8	30.3	22.8	15.5
Salable By-Product		80% H_2SO_4, 60% HNO_3				
By-Product Production (Tons/Day)	- H_2SO_4	481	971	1,446	680	388
	- HNO_3	148	148	148	104	59
By-Product Credit ($MM/Yr)	- H_2SO_4	0.67	1.36	2.02	0.95	0.54
	- HNO_3	0.65	0.65	0.65	0.45	0.26
Total Gross Production Cost ($MM/Yr)		10.45	10.51	10.63	7.99	5.32
$/Ton Sulfur Not Emitted		268.6	137.7	94.1	149.5	174.1
Mills/KWH		1.49	1.50	1.53	1.63	1.90
% Increase of Power Cost		22.1	22.2	22.7	24.1	28.1
Total Net Production Cost ($MM/Yr)		8.84	8.03	7.33	6.26	4.33
$/Ton Sulfur Not Emitted		227.1	104.7	64.6	117.0	141.6
Mills/KWH		1.26	1.15	1.05	1.28	1.54
% Increase of Power Cost		18.7	17.0	15.6	19.0	22.8

*BASE CASE

Notes:

By-Product Values
78-80% Sulfuric Acid - $6/Ton 100% Contained
98.5% Sulfuric Acid - $12/Ton
60% Nitric Acid - $25/Ton 100%
Sulfur - $20/Short Ton
Plant Load Factor - 80% (7,000 Hr. Stream Year)
Electric Power Manufacturing Cost - 6.75 Mills/KWH
Plant Location - Cincinnati, Ohio

TABLE B-12. PARAMETRIC STUDY
MOLTEN CARBONATE PROCESS

PLANT SIZE	1,000 MW 2% S	1,000 MW 3.5% S*	1,000 MW 5% S	700 MW 3.5% S	400 MW 3.5% S
Percent of Sulfur Removed from Flue Gas	91	95	96.5	95	95
Fixed Capital Investment ($MM)	31.4	34.6	37.0	27.2	18.2
Salable By-Product			SULFUR		
By-Product Production (Tons/Day)	144	261	380	183	104.4
By-Product Credit ($MM/Yr)	0.84	1.52	2.22	1.07	0.609
Total Gross Production Cost ($MM/Yr)	10.68	12.32	13.74	9.46	6.25
$/Ton Sulfur Not Emitted	234.5	148.4	114.0	162.8	138.3
Mills/KWH	1.53	1.76	1.96	1.93	2.23
% Increase of Power Cost	22.7	26.1	29.0	28.6	33.0
Total Net Production Cost ($MM/Yr)	9.84	10.8	11.53	8.40	5.64
$/Ton Sulfur Not Emitted	216.1	130.1	95.6	144.5	170.0
Mills/KWH	1.41	1.54	1.65	1.71	2.02
% Increase of Power Cost	20.9	22.8	24.4	25.3	29.9

*BASE CASE

Notes:

By-Product Values
78-80% Sulfuric Acid - $6/Ton 100% Contained
98.5% Sulfuric Acid - $12/Ton
60% Nitric Acid - $25/Ton 100%
Sulfur - $20/Short Ton

Plant Load Factor - 80% (7,000 Hr. Stream Year)
Electric Power Manufacturing Cost - 6.75 Mills/KWH
Plant Location - Cincinnati, Ohio